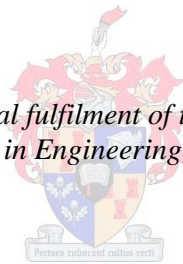


An investigation into control techniques for cascaded
plants with buffering, to minimise the influence of process
disturbances and to maximise the process yield

by
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*Thesis presented in partial fulfilment of the requirements for the degree
of Master of Science in Engineering at Stellenbosch University*



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Declaration

By submitting this thesis electronically, I declare that the entirety of the work contained therein is my own, original work, that I am the owner of the copyright thereof (unless to the extent explicitly otherwise stated) and that I have not previously in its entirety or in part submitted it for obtaining any qualification.

Date: November 2010

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Abstract

The Coal to Liquid facility, Sasol, Secunda operates as a train of processes. Disturbances and capacity restrictions can occur throughout the plant and the throughput fluctuates whenever disturbances occur. When capacity restrictions occur in a sub-plant and more substances enter the sub-plant than can be processed, the extra substances are flared or dumped and therefore lost. To reduce losses and extra costs and to maximise the throughput of the whole plant, supervisory control is implemented over the whole plant system.

Each process in the process train is controlled with regulatory controllers and the overall process is then controlled with a supervisory controller. These two sets of controllers operate in two different layers of control, with the regulatory controllers the faster inner layer. The supervisory control is the outer layer of the two control layers. The supervisory controller takes over the work of the human operator by deciding on the changes in total throughput as well as the set points for each individual process. These set points for each process are then followed with the regulatory controllers. For the regulatory control of the system, different control methods are investigated and compared. The different control methods that are looked at are PI control, Linearised State Feedback control, Fuzzy Logic control and Model Reference Adaptive Control.

After an investigation into the various control methods Fuzzy Logic control was chosen for the regulatory as well as the supervisory control levels. Fuzzy Logic control is a rule based control method. Fuzzy variables are everyday terms such as *very slow* or *nearly full*. These terms are easy to understand by the operator and multi-variable control is possible with Fuzzy Logic control without an accurate mathematical representation of the system. These facts made Fuzzy Logic control ideal for this implementation.

To improve the profit of the Coal to Liquid facility the throughput was maximised. The combination of regulatory and supervisory controllers minimised losses and rejected disturbances. This resulted in a smoother output with maximum profit.

Opsomming

Die Steenkool-na-Olie fasiliteit, Sasol, Secunda funksioneer as 'n trein van prosesse. Versteurings en kapasiteit beperkings kan deur die hele aanleg voorkom en die deurset wissel voortdurend wanneer versteurings voorkom. Wanneer kapasiteit beperkings voorkom in 'n aanleg en meer stowwe word in die aanleg ingestuur as wat dit kan verwerk, word die ekstra stowwe gestort en dit gaan verlore. Om verliese en kostes te verminder en om die deurset van die hele aanleg te vergroot, is oorhoofse beheer geïmplementeer oor die hele stelsel.

Elke proses in die trein van chemiese prosesse word beheer met regulerende beheerders. Die totale proses word dan beheer met 'n oorhoofse beheerder. Hierdie twee tipes beheerders funksioneer in twee lae van beheer met die regulerende beheerders die vinniger binneste laag. Die oorhoofse beheerder vorm die buitenste laag van die twee beheer lae en neem die werk van die menslike operateur oor deur die veranderinge in die totale deurset, sowel as die stelpunte vir elke afsonderlike proses, te bepaal. Hierdie stelpunte vir elke proses word dan met die regulerende beheerders gevolg. Verskillende beheer metodes is ondersoek vir die regulerende beheer van die stelsel. Die verskillende beheer metodes waarna gekyk word, is PI beheer, Geliniariseerde Toestands Terugvoer beheer, Wasige Logiese beheer en Model Verwysing Aanpassende beheer.

Na 'n ondersoek na die verskillende beheer metodes is Wasige Logiese beheer gekies vir die regulerende asook die oorhoofse beheer. Wasige Logiese beheer is 'n reël gebaseerde beheer metode. Wasige Logika veranderlikes is alledaagse terme soos *baie stadig* of *byna vol*. Hierdie terme is maklik om te verstaan deur die operateur. Meervoudige-veranderlike beheer is moontlik met Wasige Logiese beheer sonder 'n akkurate wiskundige voorstelling van die stelsel. Hierdie feite maak Wasige Logiese beheer ideaal vir hierdie doel.

Om die wins van die Steenkool-na-Olie fasiliteit te verbeter, is die deurset gemaksimeer. Die kombinasie van regulerende- en toesighoudende beheerders beperk verliese en verwerp versteurings. Dit lei tot 'n gladder uitset en 'n maksimum wins.

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Nomenclature

Abbreviations and Acronyms

MRAC	Model Reference Adaptive Control
PID	Proportional Integral and Derivative (Control)
PI	Proportional and Integral (Control)
SP	Set Point
CR	Capacity Restriction
VP	Valve Position

Greek Letters

α	Valve characteristic
ω_n	Natural frequency
τ	Time constant
ρ	Water density
ζ	Damping ratio
θ	Temperature
ϵ	Control Parameter with Model Reference Adaptive Control

Lowercase Letters

a	$\frac{1}{\tau}$ with transfer function $G(s) = \frac{B}{\tau s + 1} = \frac{b}{s + a}$
b	$\frac{B}{\tau}$ with transfer function $G(s) = \frac{B}{\tau s + 1} = \frac{b}{s + a}$
c	Specific heat of a medium
g	Gravitational acceleration
$h(t)$	Height at time step t
k	Used for different gains, should be specified when used
$p(t)$	Pressure at time step t
$q(t)$	Flow rate at time step t

Uppercase Letters

A	Base area of tank
B	Gain of transfer function $G(s) = \frac{B}{\tau s + 1}$, used in height and flow control
C	Capacitance
H	Height of liquid level in tank
M	Mass of substance
Q	Flow rate
R	Resistance
V	Volume of tank
W	Heater power

Chapter 1

Introduction

This thesis is an investigation into various control techniques for cascaded plants with buffering. The goals are to minimise the influence of process disturbances and to maximise the process yield at the output.

The Sasol Secunda factory is a Coal to Liquid production facility [3]. A large portion of the facility consists of several gas processing units (sub-plants) configured in series. Each unit in turn consists of identical sub-equipment (referred to as trains) which are connected in parallel. There are no hold-up facilities between the processing sub-plants, with the exception of the inter-connecting pipe work between the units. This requires that the throughput rates of individual units have to be co-ordinated effectively in real time to maintain the overall material balance of the facility. An indication of a closed material balance is a stable pressure in all the interconnecting lines. The basic control philosophy to maintain the material balance is to set the production rate of one of the units and adjust the rates of the others accordingly. This is basically done by pressure feedback control on each interconnecting header.

The processing capacity of an individual unit may become constrained at some point. It is usually related to trips and breakages of a process train but it can also be related to other process constraints. When this happens, a knock-on effect is seen on the up stream and downstream equipment. During such an event, a temporary over-production situation develops on upstream units. Some of this can be rectified by flaring (dumping) the product in order to create an artificial consumer.

The limitations of distributed feedback control are often manifested in the following:

- Delayed reaction to a disturbance, leading to sub-optimal control of material balance leading to further production losses. A further effect of this is often dynamic over-compensation to restore material balance. Some units have a slow production ramp-up rate and unnecessary over-shoot of control action takes time to correct, which lead to sub-optimal production rates.
- Reaction to minor frequent disturbances causing frequent small adjustments to unit production rates which often have to reverse from minute to minute. For

these small changes the surge capacity in the interconnecting header can be utilised better.

The result of sub-optimal automatic coordination of the unit production rates is that higher variance is distributed across the critical process variables of the factory. This variance, in turn, is met by a conservative production rate of the facility, to avoid violations of pre-determined values for these variables. This production rate is manually adjusted from time to time in an attempt to maximise the production rate of the facility.

Two opportunities exist for improving on the basic control system. Firstly, to better co-ordinate the production rates of units dynamically to reduce variance in critical process variables and then secondly to automatically set the factory throughput based on the prevailing constraints. More advanced and centralised control strategies should be investigated to achieve this.

In this thesis, only one of the production trains is looked at. This train of chemical processes is represented by a cascaded system of sub-plants, connected via tanks, through which a liquid flows.

At first a simple representation of the system is used. This representation consists of only two tanks, connected through a valve at the bottom. Different control methods are used to control the outputs. The different control methods are then evaluated under different circumstances and compared.

Once this is done, the representation of the system is changed to be a more realistic and more complex one. The cascaded sub-plants with buffering are simulated by four different first order plants with time delays, connected in series, with three small accumulator tanks in between to act as buffers. The third sub-plant has a non-linear process gain, while the other three sub-plants are linear. Supervisory control is necessary to control the overall plant system. Different control methods are implemented and compared. These results are used to conclude which control methods are the best for this process.

1.1 Problem Description

The problems that will be addressed throughout this thesis are based on a Coal to Liquid production facility. The whole plant system will be represented by a few cascaded sub-plants with buffering. Whenever minor frequent process disturbances occurred at the input, the disturbances were visible throughout the whole plant system. Capacity restrictions throughout the plant system decreased the total process yield. These capacity restrictions can also cause losses through dumping and flaring of up-stream products. The goal will be to reject disturbances as well as to maximise the process yield and minimise the losses. The representation of this process, as used in this thesis, is shown in Figure 1.1. The sub-plants are connected via buffering tanks and the different gases and other substances are represented by a liquid flowing through the tanks.

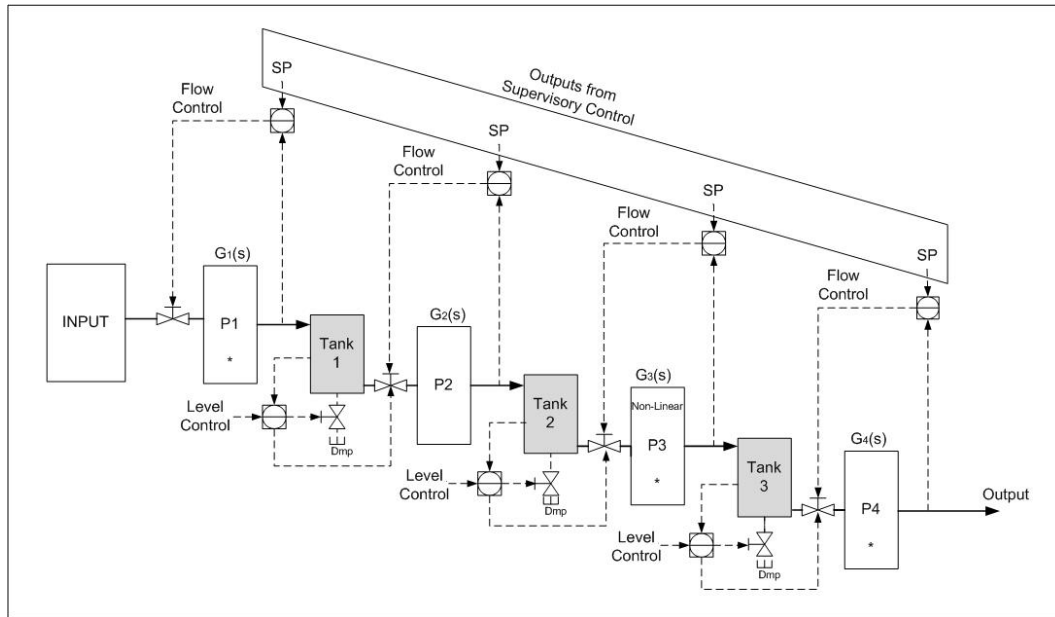


Figure 1.1 – Multiple Tank System

1.1.1 Process Description

- Each sub-plant has an associated capacity and load setting, with a maximum capacity of 100% of maximum flow. Capacity restrictions can occur in the input as well as in the first, third and fourth sub-plants. Then their capacities are less than 100% of maximum flow. No capacity restriction will occur in the second sub-plant.
- All the sub-plants have linear first order transfer functions, except the third sub-plant which has a non linear gain. In each sub-plant a dead time of about 10% - 15% of the time constant, τ , exists. One sub-plant should have a larger dead time of approximately the value of τ .
- The non-linear process (third sub-plant) has a constant dead time and τ , but smaller process gain (up to 25%) at lower inputs.
- Buffers have dumping valves which are activated if the height of the liquid in the buffer exceeds 80%. The dumped liquid will be lost and dumping should therefore be prevented.
- The buffers have a capacity of 100% of maximum height. The normal height values for the buffers are ideally at 50% of the maximum height. This will change in order to absorb the process disturbances. Still, the buffers are limited to prevent a buffer from running empty or from overflowing. These limits should keep the heights ideally between 40% and 60% of maximum height. Outside of these limits, action should take place to prevent the height from going too low or too high. When the height reaches the value of 20% of maximum height, or below that, a cut back in throughput should be activated. When the height reaches the value of 80% of maximum height, or above, the dumping valves are activated and product will be lost.

1.1.2 Control Objectives

- Honour the limits of the buffers, described in section 1.1.1
- Maximise the throughput and the total process yield
- Minimise the loss of product through dumping valves

1.1.3 Definition Of Performance Measurement Criteria

It is necessary to design a baseline control system. All the other control methods are then compared to the baseline control system.

- Disturbances are added to the process and then different controllers are compared.
- The following performance measurement standards are used:
 1. Frequency analyses of flow: Disturbances, with different frequency components, are added to the process, then the outputs are measured and compared.
 2. Statistical analysis of flow and height: The standard deviation of flows and levels are measured.

1.1.4 Disturbances

The disturbances that can occur throughout the system are listed below:

- Sustained step disturbance
- Temporary step disturbance (Pulse disturbance)
- Oscillatory disturbance (Period \ll dominant Plant τ)
- Reduction in input capacity as well as in capacities of sub-plant 1 (P1), sub-plant 3 (P3) and sub-plant 4 (P4).

1.2 Chapter Overview

Chapter 1 gives an introduction to the problem as well as an overview of the whole thesis.

Chapter 2 is a literature study on various topics covered in the thesis. The background of different control methods and other control techniques used in the thesis, are described here. Previous work published on the control problem is discussed to describe its influence on this thesis.

Chapter 3 is used to describe the first simplified representation of the cascaded system. Models were first built to represent a single tank. These models of tanks could be connected to each other to form a chain of tanks in series. The simplified representation consists of only two tanks in series. Three different control methods were implemented on this system and they are: Linearised feedback control, Fuzzy Logic control and Model Reference Adaptive Control (MRAC). The results from the simulations done on this system, as well as the models built for each individual tank, were used in the next implementation, where a more accurate representation of the system is controlled.

Chapter 4 gives an overview of the more accurate representation of the system described in section 1.1. Initially, the heights of the liquid in the buffers were controlled at the nominal height value of 50%. This was done to make it possible to control the buffer capacity. The existing buffer capacity of the actual gasification plant is very small and the flow throughout the whole plant system, without using sufficient buffers, was examined. Three different control methods were used and compared. They are Proportional and Integral (PI) control, Fuzzy Logic control and MRAC. From these results it could be seen that the buffer can be controlled. Without the use of the buffers, the disturbances that occurred could not be rejected sufficiently. To make proper use of the buffers, the output flow from the sub-plants could be controlled. This will be covered in the next chapters.

Chapter 5 describes how the flow rates from the various sub-plants are controlled. From the results in chapter 4, it was concluded that the buffers should be used to absorb the disturbances in the process. To make use of the buffer capacities, but also to acknowledge the restrictions on the buffers, new controllers were designed. The flow rates of a liquid from the sub-plants were now controlled. This was done by valves between the buffers and the sub-plants, which means the inlet flow was controlled to achieve the correct outlet flow.

The three different control methods used in chapter 4, were used for the flow control as well. To control the flow rates, each sub-plant has its own controller and should follow its own set point. These set points depend on the maximum throughput as well as the restrictions that can occur in the different sub-plants. To determine these set points, a supervisory controller was designed.

Chapter 6 describes the supervisory control, used to determine set point values for the controllers, described in chapter 5. The supervisory controller is a controller that gets information from the whole plant system and uses that information to decide what each of the regulatory controllers should achieve. The supervisory controller will take over the work of the human operator.

Chapter 7 gives a conclusion on how the problem is solved. Recommendations for further research and practical implementation are done here.

Chapter 2

Literature Study

This chapter offers an overview of the literature used to get a background of the various topics which contributed to this thesis.

The main objective of the thesis is to investigate different control techniques for cascaded plants with buffering. Therefore different control techniques and their characteristics will be investigated. Other techniques used in previous work, to control cascaded plants, are also discussed.

2.1 PID Control

The first control method that is discussed is common and widely used and known as the PID controller. The abbreviation stands for the three terms of the controller, the *Proportional*, *Integral* and *Derivative* terms.

PID control is a feedback control method, through which the output is measured against a set point and the difference is known as the error signal. The error signal is then used to determine the control signal. The design of a PID controller is generic, but each controller should be tuned to the specific system. The first of the three parameters to be tuned are the proportional gain, which gives a reaction to the current error. The second parameter is the integral gain, which gives a reaction on the integral over time of the current errors and the last parameter is the differential gain, which gives a reaction on the rate of change in the error.

The equation for the control signal in the time domain is given in equation 2.1, and the transfer function of a PID controller is given by equation 2.2 [4].

$$u(t) = K_P e(t) + K_I \int e(t) dt + K_D \frac{de(t)}{dt} \quad (2.1)$$

$$G_c(s) = K \left(1 + \frac{1}{T_i s} + T_d s \right) \quad (2.2)$$

Different variations of the controller can be used, for instance the derivative gain, T_d can be set to zero and then it is known as a PI controller. The integral gain, T_i can be set to infinity to have a PD controller.

Figure 2.1 shows a block diagram of a PID controller. $K_p = K$, $K_i = \frac{K}{T_i}$ and $K_d = KT_d$, from equation 2.2.

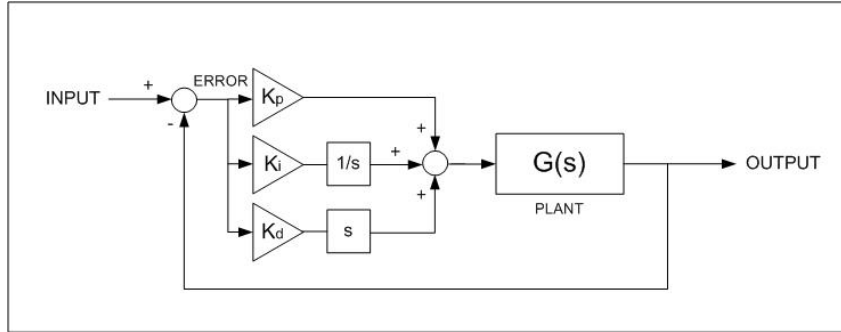


Figure 2.1 – Block Diagram Of A PID Controller

A variation of the PID controller, a PI controller, is used for base case control. PI control is sufficient for the control of slow processes. The derivative term may amplify noise. In this process, the valves and equipment can be noisy. The derivative term is not necessary, because it is a slow process. PI control is expected to be sufficient. Other control methods used are compared to this base case controller.

2.2 Fuzzy Logic Control

Fuzzy Logic Control is a control method developed to simulate human thinking. The controller is rule-based and the rules are usually in the *if-then* format. Because not all questions can be answered as true or false, there is mostly true or likely false etc. Fuzzy Logic control uses Fuzzy sets, that accommodate everything in between true (1) and false (0) values.

Fuzzy sets were invented by Lotfi Zadeh in the mid-1960s. [5] His argument was that classes of objects in the real physical world often could not be described by precise memberships, for instance the *class of tall human beings*. We cannot draw a line which separates tall people from short people at a certain height and then define people 1cm under the line as short and 1cm above it as tall. We have medium short and medium tall people. We can have extremely short or very tall ones. To accommodate these in-between values in fuzzy sets, fuzzy membership functions are used.

A set is a selection of items that can be treated as a whole. Fuzzy sets can contain many items (members), each with a probability or a grading between 0 and 1. Membership functions are the functions that attach a grading number to each element in the universe. If an object is an absolute member of the set, it will be 1 and if it is not at all a member, it will be 0. Anything in between is also possible, therefore an item can be a partial member by assigning any real value between 0 and 1 to its grading. For instance, in a *set of long hiking distances*, 10km can have a grading of 0.8 while 3km has grading of 0.2. The elements of a Fuzzy set are taken from a universe which contains all the possible items [5][6]. The membership functions can be continuous or discrete.

In most cases continuous membership functions are used and they can be *bell shaped* (π -curve), *S shaped* (s-curve), *Reversed s shaped* (z-curve), *Triangular* or *Trapezoidal*. In the case of discrete membership functions, values in a list are used.

The fuzzy logic controller consists of three different parts. The first part is the *fuzzification* of the inputs. Measurements are converted from the numerical values from sensors or measurement equipment into fuzzy variables. The second part is the *inference system* with the *rule base*. This is where the fuzzy inputs are used to create the fuzzy outputs by means of implementing the rules. The last part is where the fuzzy outputs are again converted into a value used by the system, like a current of $4 - 20mA$ or a valve position. Figure 2.2 shows a block diagram of a simple fuzzy system.

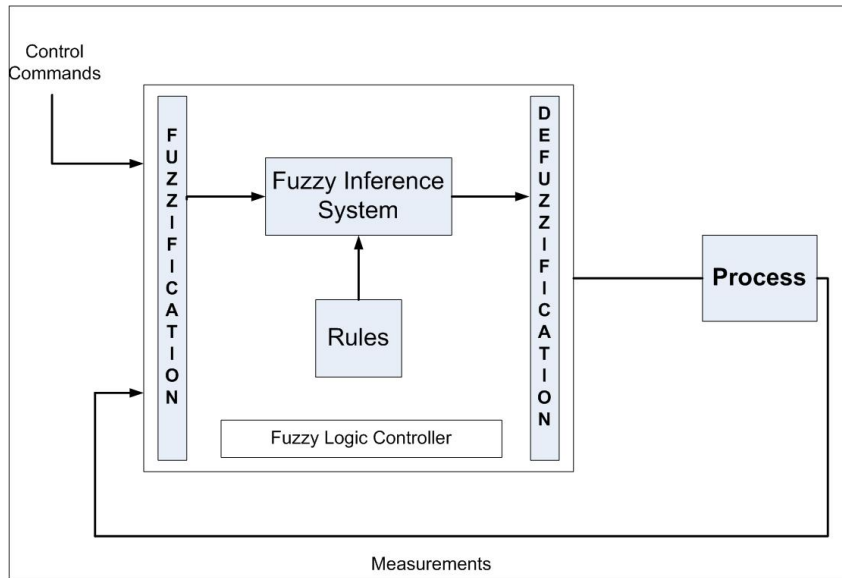


Figure 2.2 – Diagram Of A Fuzzy Logic System

The Fuzzy Inference system uses rules to simulate human thinking. The human's capacity to reason with approximations made it possible to adapt to unfamiliar situations where they could gather (sometimes subconsciously) valid information and discard irrelevant details. This information is more often than not vague, qualitative and general. Fuzzy Logic provides an inference morphology that makes human-like thinking or reasoning possible. The Fuzzy Rules are symbolically written as:

IF (premise i) THEN (consequent i)

where i is each rule in the set of rules. The input premise can be a single statement such as (IF x_1 is A), but one can also make use of the logical *AND* and *OR* to accommodate more than one statement. *AND* is used for *intersection* of two statements and *OR* is used for *union* of two statements. This is then used in the form (IF x_1 is A AND x_2 is B). Here x_1 and x_2 are inputs and A and B are fuzzy compounds. The rule can also be used with a *NOT*, for instance, (IF x_1 is A OR x_2 is NOT A). The consequence of each rule can be defined in two ways [6]. This separates the two major types of fuzzy rules.

The first type is known as Mandani fuzzy rules and the second type as Takagi-Sugeno rules.

Mandani Fuzzy Rules are rules of the form

$$\text{IF } (x_1 \text{ is A}) \text{ AND } (x_2 \text{ is B}) \text{ THEN } (y_1 \text{ is C})$$

where A, B and C are fuzzy values.

Takagi-Sugeno Fuzzy Rules are rules of the form

$$\text{IF } (x_1 \text{ is A}) \text{ AND } (x_2 \text{ is B}) \text{ THEN } (u = f(x_1, x_2))$$

both A and B are fuzzy values and u is a function of the input variables.

The TILShell product [5] makes a few recommendations on where to start with the design of the membership functions. The first is to start with triangular sets and to choose three sets per variable. The membership functions for a specific input or output is initially chosen as identical triangles of the same width. Each value of the universe should be a member of at least two sets. The rules will be applied so that more than one rule can be applied to each element. This will make the control smoother. These recommendations were used and then the membership functions were adapted until the performance requirements were satisfied.

There are many reasons for considering Fuzzy Logic control. [5] The first reason is that multiple inputs and multiple outputs can easily be controlled without theoretical difficulties. In this case it is a distinct advantage, because different inputs like flow rates and height values of different sub-plants and tanks can all be used in the same controller to calculate different outputs. The second advantage is the fact that the process model is not needed. Therefore, no uncertainties or approximations in the process model will have an influence on the performance of the controller. The third reason to use Fuzzy Logic control is because of the fact that everyday terms are used in the rule base. These *if-then* rules can be understood by any operator without computing skills. It is easy to understand the rules and therefore problems can easily be addressed without looking at mathematical models. The Fuzzy Logic controller was compared with the PID controller in simulations and experiments. [5] The Fuzzy Logic controller often showed more robustness, slower rise time, faster settling time and less overshoot. The control signal was also often much smoother. Therefore, although Fuzzy Logic control involves building rather arbitrary curves of fuzzy sets and requires knowledge of fuzzy set theory, it holds many advantages for this application.

2.3 Model Reference Adaptive Control

Model Reference Adaptive Control is an adaptive control method that compares the output of the plant that needs to be controlled, to a chosen reference model and then uses the error in the output to change the controller parameters. To adjust the parameters, two different methods could be used. These are the use of a gradient method or by applying stability theory. The original solution for MRAC was developed at the

Instrumentation Laboratory at MIT, and is known as the MIT rule. This is a gradient approach and the method used in this thesis. To look at the stability theory, Lyapunov's Stability Theory could be used [1], paragraph 5.4. A block diagram of a MRAC is shown in Figure 2.3.

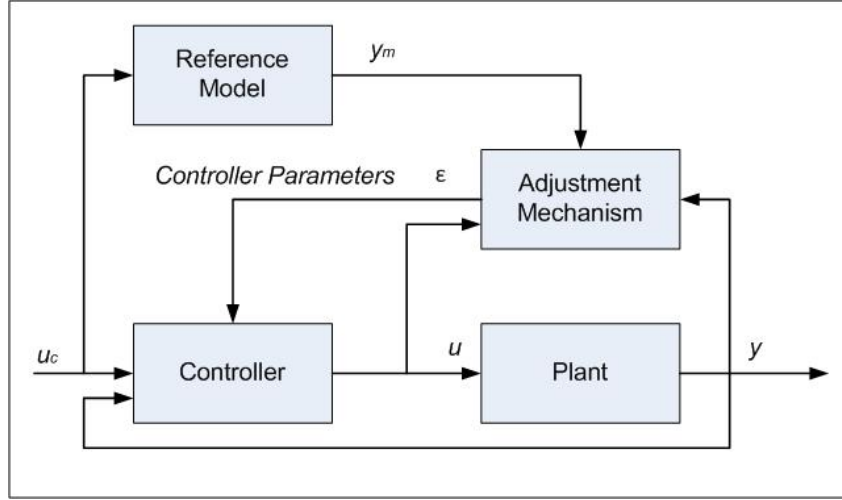


Figure 2.3 – Block Diagram Of A Model Reference Adaptive System, [1]

To determine the gain functions, the MIT rule is used. This rule states that $\frac{d\epsilon}{dt} = -\gamma \frac{\partial J}{\partial \epsilon}$ where J is the quadratic error cost function, γ is positive gain and ϵ is the adjustment signal. Then $\frac{d\epsilon}{dt} = -\gamma \frac{\partial e}{\partial \epsilon} e$, with e the error signal. The cost function is then minimised and the error will go to 0. The partial derivative, $\frac{\partial e}{\partial \epsilon}$, is known as the sensitivity parameter [1].

The cascaded process, which should be controlled, consists of various non-linear relationships. The flow of water through a valve is non-linear and the gain of the third sub-plant is non-linear. This means that to implement control methods such as PID, PI or Model Reference Adaptive Control, the system should be linearised at certain work points. The controller performance will be worse at values different from these chosen work points when using a non-adaptive control method such as PI control. This is why MRAC is considered. The process will at times operate at values different from the nominal work points. In these situations the MRAC is expected to perform better than a normal PI controller because the controller gain changes when the system changes.

2.4 Cascade Control

Cascade control loops are widely used in the process control industry to control pressure, temperature and flow. It is used to improve the performance, reject disturbances or increase the controller's speed [7]. In many cases these original control loops consist of long time delays or strong disturbances. These time delays and disturbances are then dealt with by using cascade control. A secondary inner loop is added in cascade with the system. This secondary inner loop takes care of the control much faster, which means

the lag as well as the effect from disturbances can be minimised. An example of cascade control is temperature or flow control, where the sensor is at a certain distance from the point where a disturbance occurs. Normal feedback control will only make changes once the sensor measurement shows the effect of the disturbance. A secondary loop can be added, that measures the disturbance and then starts taking action before the effect is shown in the plant. A diagram of a cascade control system is shown in Figure 2.4, [7]. The same principle is used in [8], but then it is also used for parallel processes and not just for processes in series.

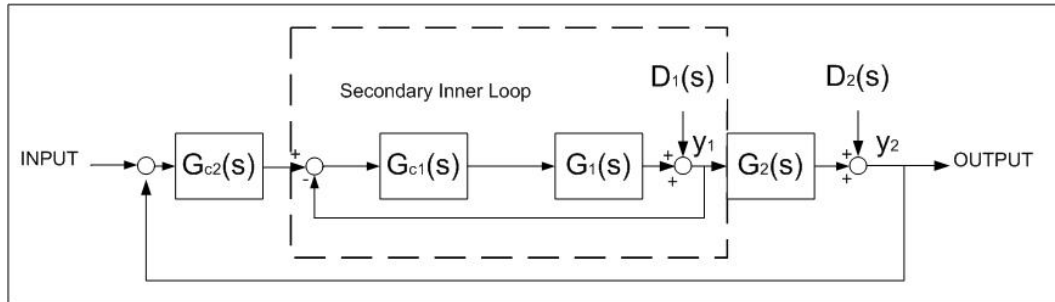


Figure 2.4 – Diagram Of A Cascade Control System

When one look at the cascaded system used in this thesis, the same principle used in cascade control, can be applied. The individual sub-plants can be controlled individually and form the secondary inner loops. These controllers are the flow rate controllers for each sub-plant or the height controllers for each buffer. They use the outputs of each individual sub-plant of buffer and not the output of the overall plant system. These inner loops operate faster than the supervisory controller of the whole system.

2.5 Supervisory Control

A Supervisory controller is designed to mimic the human operator, according to Jantzen [9]. Goals for supervisory control are safety, product quality and economic operation. These goals should be prioritised and safety gets highest priority.

In both [10] and [11], they describe three steps to plant wide control. (1) Determination of control variables, manipulated variables and process measurements, (2) control configuration and (3) controller selection. [11] describes self optimising control as the state when we can achieve an acceptable loss with constant set point values for the controlled variables, without the need to re-optimize when disturbances occur. The article refers to a typical control hierarchy as in Figure 2.5 when discussing self optimising control. Step one, where the control variables, manipulated variables and process measurements are determined, is a crucial step to the successful design of a self optimising controller. The best set of control variables are selected by minimising a loss function.

In this optimisation problem, one has three choices:

- (1) Open-loop implementation (cannot reject disturbances)
- (2) Closed loop implementation with separate control layer
- (3) Integrated optimisation and control (very complicated)

The second possibility, a closed loop implementation with a separate control layer, will be used. This means that the optimiser is outside the controller. The optimiser determines the set points of the controlled variables of the regulatory controller. The regulatory controller then controls the process according to these inputs from the optimiser.

In [12] the same principle is implemented on a ball mill grinding circuit. Here the different controlled variables (ore feed rate, feed water rate and the sump water rate) are under normal PID control. The set points of all these controllers come from the supervisory level.

In [2], a study is done on the production chain of Statoil Hydro's Snohrit plant in Hammerfest, Norway. When each part of a process or value chain is optimised individually, a dividing wall between parts exists, even though the different parts are tightly connected. This leads to poor optimisation of the whole process. Model-based optimisation is used to find optimal operation when unexpected operational events are present. The study starts by setting up a process control hierarchy for the problem. This control hierarchy can be seen in Figure 2.5. In the study, focus is placed more on Scheduling and Site-wide Optimisation, while the objective of this thesis is to focus more on Supervisory and Regulatory Control (Control layers).

In [13], the objective was to maintain plant operation near optimum, even with disturbances and other external changes. Typical Real Time Optimisation (RTO) is model-based and implemented on top of unit-based multi variable controllers. The RTO layer is between the Production planning (Scheduling) and Local Controller layers. Conventionally, steady-state model based RTO formulation was used. Most integrated plants have very long transient dynamics. This limited the frequency of optimisation because plants would seldom be in steady state. The reason is that additional changes would occur in the meantime. Once a change has occurred, it could take a long time to reach a new steady state. Also, optimal operating conditions calculated at steady-state may be suboptimal or even infeasible. This is due to disturbances, model errors, unit interaction and transient dynamics. The steady state assumption precludes the use of dynamic degrees of freedom available in the plant (e.g. storage capacities). To overcome the steady state drawbacks, a RTO slower than the local unit MPCs, is suggested. This method performs an RTO at a lower frequency than the MPC (Model Predictive Controller) frequency, but it does not have to wait for steady state to be reached. Dynamic optimisation when a slow-scale model is used, is described. A few examples are given and their results are discussed. Through the examples it is shown that a slow-scale model can provide an efficient RTO solution.

The supervisory controller will have an overall view of the whole plant system and will have access to all the information of the different parts of the plant. This information

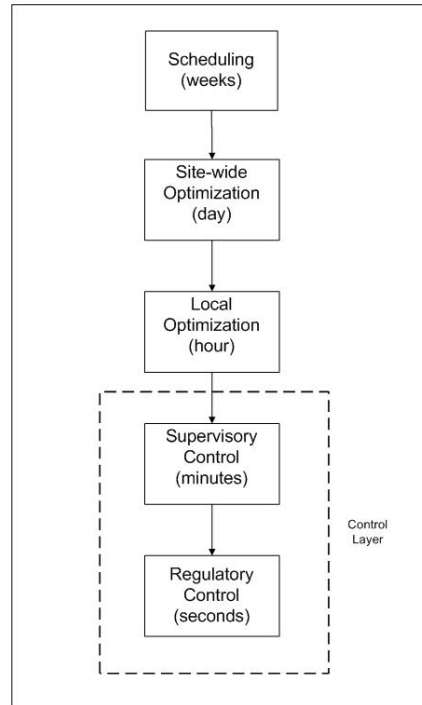


Figure 2.5 – Control Hierarchy [2]

will be used to calculate the best set points for the regulatory controllers. The regulatory controllers will only have access to information on the specific area of the plant and will control the outputs according to the set points from the supervisory controller.

2.6 Conclusion

The different control methods discussed in this literature overview, were found suitable for investigation for the cascaded control, because of different reasons.

PID control is widely used and understood in the process control industry. This will make the PID controller a suitable control technique for baseline control. This controller can then be compared to the other, less common control methods. The PID controller has a generic form and the tuning gains can be calculated for the specific system.

Fuzzy Logic control makes multiple variable control easier. This will make the simultaneous control of flow through the sub-plants as well as the height of the buffers possible, without the need for a mathematical representation of the system. Fuzzy Logic control will be effective for a supervisory controller, where many different inputs are considered.

Model Reference Adaptive Control will be investigated to see whether it will improve the controller performance of non-linear systems. Non-linear systems can be controlled by linearizing the system at a certain work point. When the operation takes place at values not equal to the chosen work points, control performance can decrease. With

a controller that adapts its controller gains with changes in the system, better results with non-linear systems can be expected.

Supervisory control will be applied on the overall system. This control layer will be used to control the whole plant system by calculating efficient set point values for the regulatory controllers.

This literature study has provided the required background information. The information was used and some methods were implemented to design the controllers for the cascaded plant system. Other literature used are [14], [15], [16], [17], [18], [19], [20], [21] and [22].

Chapter 3

Non-Linear Tank Model

The cascaded plant representation, described in section 1.1, consists of various sub-plants, connected in series. There are buffers between these tanks. These buffers are tanks with a certain capacity. To create a model of the whole cascaded plant system with buffer tanks, a model of a single tank should first be created. In this chapter such a model is created and then a simplified representation of the cascaded plant system is introduced. This simplified representation consists of only two tanks, connected by a valve at the bottom. The output flow rate and temperature are controlled using various control methods. These control methods are then implemented on the more realistic plant representation, and will be discussed in the chapters to follow.

3.1 Model Design

A model is developed to simulate a tank which can be connected to other plants or tanks in a chain, which then represents a chain of different chemical processes. The model is based on a tank of which the base area (A) can be chosen. Other parameters such as initial height (H_0) and initial temperature (θ_0) may also be inserted by the user. The temperature (θ_{in}) and flow rates (Q_{in} and Q_t) of the inflowing liquid are also input parameters to the model of the tank system. The tank can have inputs either through an inflow at the top (Q_{in}), or through a connection from the previous tank (Q_t). Each tank can have an outflow either to the next tank through a valve (Q_w), or an outlet to the atmosphere, through a valve (Q_{out}). In the case where the outlet of one tank is by a valve to the next, the height of the following tank (H_w) is needed. The height of the previous tank is denoted as (H_t) and the temperature of the previous tank is (θ_t). The outputs are the outflow temperature (θ_{out}), the outflow flow rate (Q_{out}) through a valve with valve position (VP) and valve characteristic (α), the flow rate to the next tank (Q_w) through a valve with valve position (VP_w) and valve characteristic (α_w), and the height (H). The design of the model of the tank is based on the following differential equations, based on the equations presented in [23], p. 98 - 121:

Flow: The following hydraulic equations are used:

$$p(t) = \rho gh(t) \quad (3.1)$$

$$C = \frac{A}{\rho g} \quad (3.2)$$

Equation 3.1 states that the pressure as a function of time, $p(t)$, is directly proportional to the height, $h(t)$. The relationship between them is given by the factor of $\rho \times g$, with ρ the density of the liquid, in this case water, and g the gravitational acceleration. From equation 3.2 it is shown that the hydraulic capacitance, C , is a function of the area of the tank, A , the water density, ρ and gravitational acceleration, g .

$$C \frac{dp(t)}{dt} = q(t)_{inflow} - q(t)_{outflow} \quad (3.3)$$

therefore

$$\begin{aligned} A \frac{dH}{dt} &= Q_{inflow} - Q_{outflow} \\ &= (Q_t + Q_{in}) - (Q_{out} + Q_w) \\ &= (Q_t + Q_{in}) - (\alpha V P \sqrt{H} + \alpha_w V P_w \sqrt{H - H_w}) \end{aligned} \quad (3.4)$$

Temperature:

$$C \frac{d\theta}{dt} = \frac{\theta_{in}(t) - \theta_{out}(t)}{R_1} + \frac{\theta_t(t) - \theta_{out}(t)}{R_2} + W(t) \quad (3.5)$$

$$\theta_{out}(s) = \frac{\frac{1}{CR_1}}{s + \frac{1}{CR_1} + \frac{1}{CR_2}} \theta_{in}(s) + \frac{\frac{1}{CR_2}}{s + \frac{1}{CR_1} + \frac{1}{CR_2}} \theta_t(s) + \frac{1}{C} W(s) \quad (3.6)$$

With the thermal capacitance

$$C = Mc = V\rho c \quad (3.7)$$

and the thermal resistance

$$R_1 = \frac{1}{Q_{in}\rho c} \quad (3.8)$$

$$R_2 = \frac{1}{Q_t\rho c} \quad (3.9)$$

and

$$g = 9.81 \text{ m/sec}^2 \quad (3.10)$$

$$\rho = 1 \text{ kg/m}^3 \quad (3.11)$$

$$c = 4186 \text{ J/kg}^\circ\text{C} \quad (3.12)$$

$$A = 1 \text{ m}^2 \quad (3.13)$$

$$V = A \times H \quad (3.14)$$

with:

g = Gravitational acceleration

ρ = Water density

c = Specific heat for water

M = Mass of the substance

W = Heater Power

V = Volume (m^3)

A = Area (m^2)

H = Height (m)

To develop different controllers the following system was used initially, before a more complex and realistic system was implemented:

Two tanks are connected in series. They are connected through a valve at the bottom. The outlet from the first tank is the input for the second tank. This system is shown in Figure 3.1.

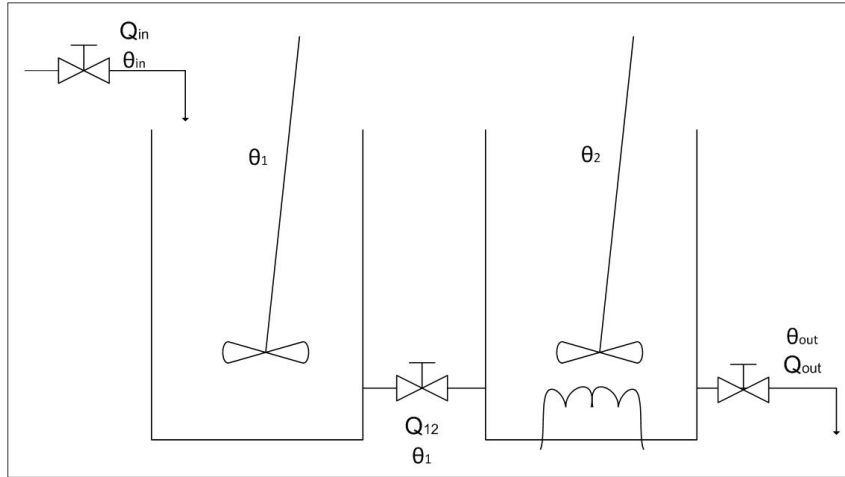


Figure 3.1 – Two Tank System

The output temperature and output flow will be controlled. This will be done with a heater in the second tank to control θ_{out} and a valve changing the input flow, Q_{in} , to control Q_{out} . The flow rate between the two tanks is Q_{12} .

3.2 Linear Control Of The Non-linear System

The described system is non-linear. It was linearised at certain work points to design controllers for the system. The work points chosen are $\bar{H}_1 = 0.301$ m and $\bar{H}_2 = 0.25$ m. These are the height values for a flow rate of 0.01 m³/sec throughout the system.

Flow: State vector, \mathbf{x}

$$\mathbf{x} = \begin{bmatrix} H_1 \\ H_2 \\ \theta_1 \\ \theta_2 \end{bmatrix} \quad (3.15)$$

Control signal, $u_1(t) = Q_{in}$

Tank 1:

$$A_1 \frac{dH_1}{dt} = Q_{in} - Q_{12} = Q_{in} - k_1 \sqrt{H_1 - H_2} \quad (3.16)$$

Tank 2:

$$A_2 \frac{dH_2}{dt} = Q_{12} - Q_{out} = k_1 \sqrt{H_1 - H_2} - k_2 \sqrt{H_2} \quad (3.17)$$

These equations are linearised at \bar{H}_1 and \bar{H}_2 to provide equal steady state responses, and not equal dynamic responses. The linearised equations are as follows:

Tank 1:

$$A_1 \frac{dH_1}{dt} = Q_{in} - k'_1 H_1 + k'_1 H_2 \quad (3.18)$$

Tank 2:

$$A_2 \frac{dH_2}{dt} = k'_1 H_1 - k'_1 H_2 - k'_2 H_2 \quad (3.19)$$

Where $k_1 = \alpha_{12} V P_{12} = 0.0443$, with valve characteristic $\alpha_{12} = 0.0443$ and $V P_{12} = 1$. Then $k'_1 = k_1 \frac{\sqrt{\bar{H}_1 - \bar{H}_2}}{\bar{H}_1 - \bar{H}_2} = 0.1962$ and $k_2 = \alpha V P = 0.02$, with valve characteristic $\alpha = 0.02$ and $V P = 1$. Then $k'_2 = k_2 \frac{\sqrt{\bar{H}_2}}{\bar{H}_2} = 0.04$.

Temperature: The thermal capacitance is given by $C = Mc = V\rho c$ with c the specific heat of a medium and M the mass of the substance.

Control signal, $u_2(t) = W(t)$.

Tank 1:

$$C_1 \frac{d\theta_1(t)}{dt} = Q_{in}(t)\rho c\theta_{in}(t) - Q_t(t)\rho c\theta_1(t) \quad (3.20)$$

$$\frac{d\theta_1(t)}{dt} = l_1\theta_{in}(t) - l_2\theta_1(t) \quad (3.21)$$

with

$$l_1 = \frac{Q_{in}(t)\rho c}{C_1} = \frac{Q_{in}}{V_1} \quad (3.22)$$

$$l_2 = \frac{Q_t(t)\rho c}{C_1} = \frac{k_1 \sqrt{H_1 - H_2}}{H_1 A_1} = \frac{Q_{12}}{V_1} \quad (3.23)$$

Tank 2:

$$C_2 \frac{d\theta_2(t)}{dt} = Q_{12}(t)\rho c\theta_1(t) - k_2\sqrt{H_2}\rho c\theta_2(t) + W(t) \quad (3.24)$$

$$\frac{d\theta_2(t)}{dt} = l_3\theta_1(t) - l_4\theta_2(t) + \frac{1}{C_2}W(t) \quad (3.25)$$

with

$$l_3 = \frac{Q_{12}(t)\rho c}{C_2} = \frac{k_1\sqrt{H_1 - H_2}}{H_2 A_2} = \frac{Q_{12}}{V_2} \quad (3.26)$$

$$l_4 = \frac{k_2\sqrt{H_2}\rho c}{C_2} = \frac{k'_2 H_2}{H_2 A_2} = \frac{k'_2}{A_2} \quad (3.27)$$

3.2.1 State Space Models

These equations were used to set up the state space for the system. The state space matrices are shown below:

Flow:

$$\begin{bmatrix} \frac{dH_1}{dt} \\ \frac{dH_2}{dt} \end{bmatrix} = \begin{bmatrix} -\frac{k'_1}{A_1} & \frac{k'_1}{A_1} \\ \frac{k'_1}{A_2} & -\frac{(k'_1 + k'_2)}{A_2} \end{bmatrix} \begin{bmatrix} H_1 \\ H_2 \end{bmatrix} + \begin{bmatrix} \frac{1}{A_1} \\ 0 \end{bmatrix} Q_{in}(t) \quad (3.28)$$

$$y_1(t) = \begin{bmatrix} 0 & k'_2 \end{bmatrix} \begin{bmatrix} H_1 \\ H_2 \end{bmatrix} = Q_{out}(t) \quad (3.29)$$

Temperature:

$$\begin{bmatrix} \frac{d\theta_1}{dt} \\ \frac{d\theta_2}{dt} \end{bmatrix} = \begin{bmatrix} -l_2 & 0 \\ l_3 & -l_4 \end{bmatrix} \begin{bmatrix} \theta_1 \\ \theta_2 \end{bmatrix} + \begin{bmatrix} 0 \\ \frac{1}{C_2} \end{bmatrix} W(t) + \begin{bmatrix} l_1 \\ 0 \end{bmatrix} \theta_{in}(t) \quad (3.30)$$

$$y_2(t) = \begin{bmatrix} 0 & 1 \end{bmatrix} \begin{bmatrix} \theta_1 \\ \theta_2 \end{bmatrix} = \theta_{out}(t) \quad (3.31)$$

To test the linear model, the following values were used:

Tank 1: $\bar{H}_1 = 0.301 \text{ m}$, $\bar{\theta}_1 = 20^\circ \text{C}$.

Tank 2: $\bar{H}_2 = 0.25 \text{ m}$, $\bar{\theta}_2 = 20^\circ \text{C}$.

Inputs: $Q_{in} = 0.011 \text{ m}^3/\text{s}$ (step up by 10% of initial flow) while the temperature is unchanged. To test the temperature model, $\theta_{in} = 15^\circ\text{C}$ (step down by 5°C), while the flow rate is unchanged.

When comparing the open-loop linearised system with the open-loop non-linear system, the following results were obtained:

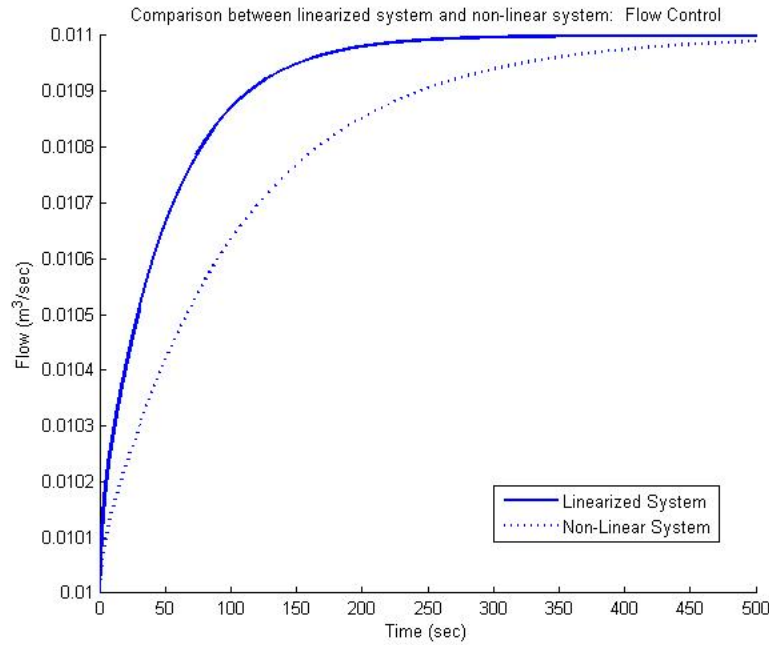


Figure 3.2 – Open Loop Comparison Between Linearised And Non-linear Systems: Flow

From Figure 3.3 it is clear that the equations, to model the temperature, are much more similar to the linearised equations. The linearisation of the flow is a bit less accurate in comparison to the non-linear system as seen in Figure 3.2. This is due to the fact that the linearised equations that describe the flow through the system, were linearised for equal steady state response and not for equal dynamic response. When the flow rate is changed, the linearised model changes and operates at values different from the values at which it is linearised. When the temperature is changed, the flow rate and therefore the values of the heights are constant and the system operates at the linearised values. The results show that when a controller is designed, the flow control will differ more than the temperature control from the linear system on which the controller will be designed.

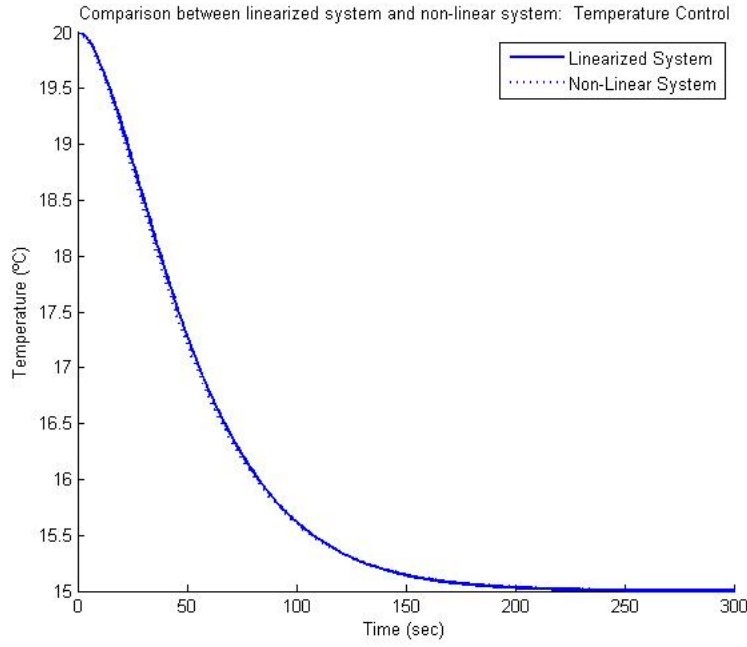


Figure 3.3 – Open Loop Comparison Between Linearised And Non-linear Systems: Temperature

3.3 Linearised Control

The linearised model of the non-linear system is used to design a controller for the non-linear system. The controller parameters are chosen for a sufficient response in flow and temperature control with the linearised model. It is subsequently implemented on the non-linear model and the results from the linearised and non-linear systems are compared.

The state space model, developed in the previous section is used. A controller with state variable feedback as well as integral action is designed. The system is augmented with the integral of the error, $e_{int}(t) = \int_0^t e(\tau) = \int_0^t y(\tau) - r(\tau) d\tau$, or $\dot{e}_{int}(t) = e(t) = \mathbf{C}x(t) - r(t)$.

The augmented system is then

$$\frac{d}{dt} \begin{bmatrix} x(t) \\ e_{int}(t) \end{bmatrix} = \begin{bmatrix} \mathbf{A} & \mathbf{0} \\ \mathbf{C} & 0 \end{bmatrix} \begin{bmatrix} x(t) \\ e_{int}(t) \end{bmatrix} + \begin{bmatrix} \mathbf{B} \\ 0 \end{bmatrix} u(t) + \begin{bmatrix} 0 \\ -1 \end{bmatrix} r(t) \quad (3.32)$$

$$y(t) = [\mathbf{C} \quad 0] \begin{bmatrix} x(t) \\ e_{int}(t) \end{bmatrix} \quad (3.33)$$

The control inputs are given by: [4], p. 694.

$$U_Q(s) = \frac{-k_{iQ}}{s} E(s) - \mathbf{k}_Q \mathbf{x} \quad (3.34)$$

$$U_\theta(s) = \frac{-k_{i\theta}}{s} E(s) - \mathbf{k}_\theta \mathbf{x} \quad (3.35)$$

and the corresponding block diagrams are shown in Figures 3.4 and 3.5.

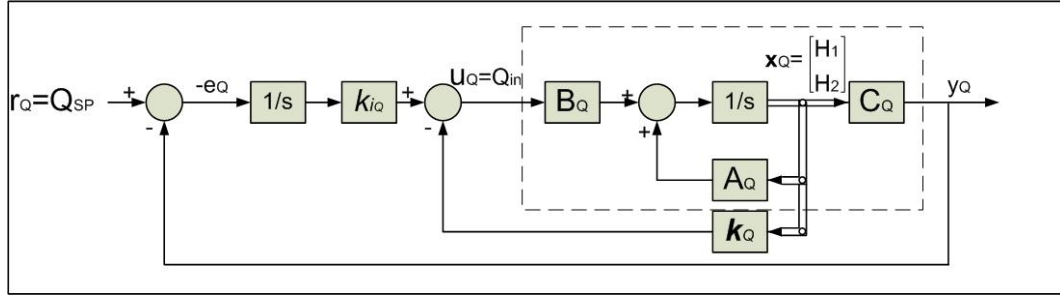


Figure 3.4 – Flow Controller For Linearised System

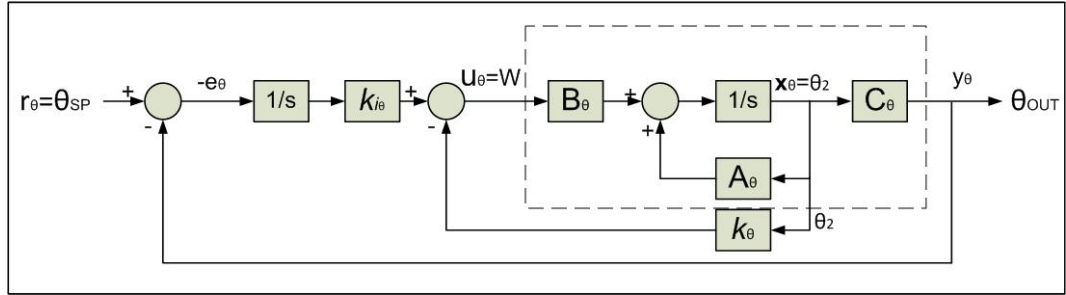


Figure 3.5 – Temperature Controller For Linearised System

The augmented flow system, from Figure 3.4 and equations 3.28 and 3.29, is given in equations 3.36 and 3.37.

$$\begin{bmatrix} \frac{dH_1}{dt} \\ \frac{dH_2}{dt} \\ \frac{de_{int}}{dt} \end{bmatrix} = \begin{bmatrix} -0.1962 & 0.1962 & 0 \\ 0.1962 & -0.2362 & 0 \\ 0 & 0.04 & 0 \end{bmatrix} \begin{bmatrix} H_1 \\ H_2 \\ e_{int} \end{bmatrix} + \begin{bmatrix} 1 \\ 0 \\ 0 \end{bmatrix} Q_{in}(t) \quad (3.36)$$

$$y_1(t) = \begin{bmatrix} 0 & 0.04 & 0 \end{bmatrix} \begin{bmatrix} H_1 \\ H_2 \\ e_{int} \end{bmatrix} = Q_{out}(t) \quad (3.37)$$

For the temperature control, only the temperature of the second tank is controlled. From equations 3.30 and 3.31, new state equations for $\frac{d\theta_2}{dt}$ are developed. The state equations are shown in equations 3.38 and 3.39 and the augmented state equations, from Figure 3.5 are given in equations 3.40 and 3.41.

$$\begin{bmatrix} \frac{d\theta_2}{dt} \end{bmatrix} = \begin{bmatrix} -0.04 \end{bmatrix} \begin{bmatrix} \theta_2 \end{bmatrix} + \begin{bmatrix} 0.0009556 \end{bmatrix} W(t) \quad (3.38)$$

$$y_2(t) = \begin{bmatrix} 1 \end{bmatrix} \begin{bmatrix} \theta_2 \end{bmatrix} = \theta_{out}(t) \quad (3.39)$$

$$\begin{bmatrix} \frac{d\theta_2}{dt} \\ \frac{de_{int}}{dt} \end{bmatrix} = \begin{bmatrix} -0.04 & 0 \\ 1 & 0 \end{bmatrix} \begin{bmatrix} \theta_2 \\ e_{int} \end{bmatrix} + \begin{bmatrix} 0.0009556 \\ 0 \end{bmatrix} W(t) \quad (3.40)$$

$$y_2(t) = \begin{bmatrix} 1 & 0 \end{bmatrix} \begin{bmatrix} \theta_2 \\ e_{int} \end{bmatrix} = \theta_{out}(t) \quad (3.41)$$

The open loop poles for the augmented flow transfer function are $[-0.4134, -0.019, 0]$. The flow controller is then designed for desired poles at $[-0.1 - 0.1i, -0.1 + 0.1i, -0.4134]$, which give a natural frequency of $\omega_n = 0.1414$ rad/sec and a damping ratio of $\zeta = 0.707$. For the temperature control, the open loop poles are at $[-0.04, 0]$. The closed loop poles are then chosen at $[-0.04 - 0.04i, -0.04 + 0.04i]$.

The gain vectors, $\mathbf{k}_{Q(tot)}$ and $\mathbf{k}_{\theta(tot)}$ are

$$\mathbf{k}_{Q(tot)} = [\mathbf{k}_Q \quad k_{iQ}] = [0.181 \quad 0.2654 \quad 1.0535] \quad (3.42)$$

$$\mathbf{k}_{\theta(tot)} = [k_\theta \quad k_{i\theta}] = [41.86 \quad 3.34] \quad (3.43)$$

These controllers were implemented on both the linearised and the non-linear systems. The results to follow set point values of $0.015 \text{ m}^3/\text{sec}$ for flow and 30°C for temperature are shown in Figure 3.6 and Figure 3.7.

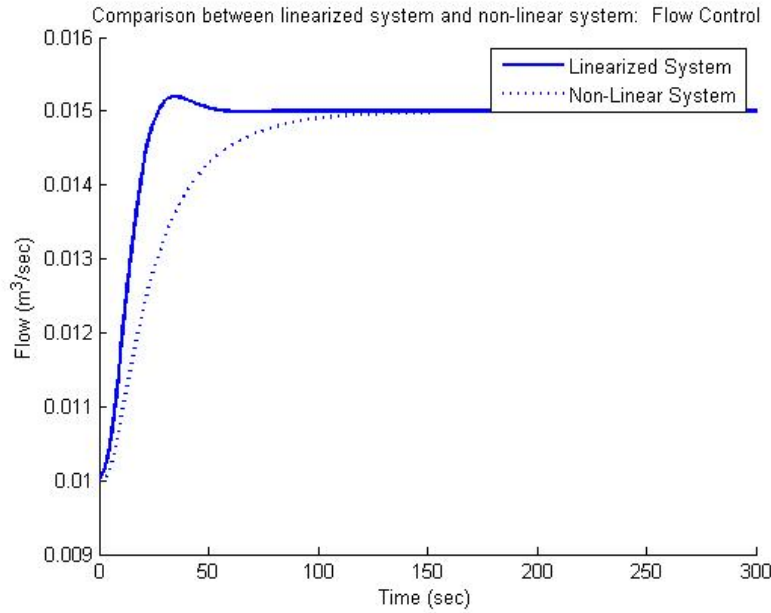


Figure 3.6 – Closed Loop Comparison Between Linearised And Non-linear Systems: Flow

The controlled temperatures of the non-linear and linearised systems are far more similar (Figure 3.7) than the controlled flow rates (Figure 3.6).

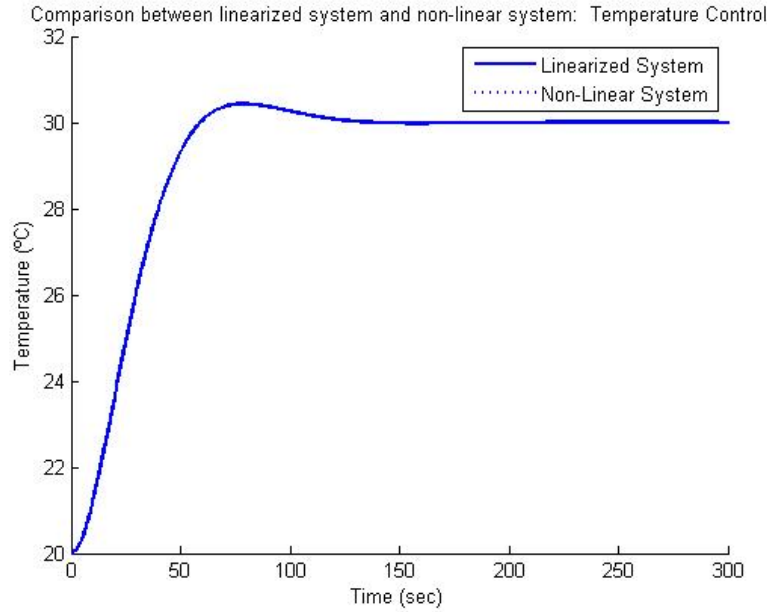


Figure 3.7 – Closed Loop Comparison Between Linearised And Non-linear Systems: Temperature

The responses to disturbances in the input flow and input temperature are investigated and can be seen in section 3.6 where it is compared to both the Fuzzy Logic controllers and MRAC, which will be discussed next.

3.4 Fuzzy Logic Control

The following paragraph describes the design of a fuzzy logic controller set with two fuzzy logic controllers. One for the control of the output temperature, θ_{out} and one for the control of the output flow, Q_{out} .

Fuzzy Logic Control is a control method developed to simulate human thinking. See section 2.2 on Fuzzy Logic Control.

3.4.1 Fuzzy Logic: Flow Control

The inputs to the flow controller is the error in the output flow, eQ_o as well as the change in error in the output flow, ceQ_o . The output is Q_{in} , the flow rate of the input stream. The membership functions are chosen as follows:

Input 1, eQ_o : The error in the output flow rate is the difference between the set point value of the flow rate and the actual measured flow rate from the tank. Five membership functions are used, labeled *Negative* (N), *Small Negative* (SN), *Zero* (Z), *Small Positive* (SP) and *Positive* (P). The range is [-1 1]. Triangular membership functions are used. Their specifications are: N is [-1.5 -1 -0.5], SN is [-1 -0.5 0], Z is [-0.5 0 0.5], SP is [0 0.5 1] and P is [0.5 1 1.5]. Figure 3.8 shows the Membership Functions of Input 1 for Flow Control.

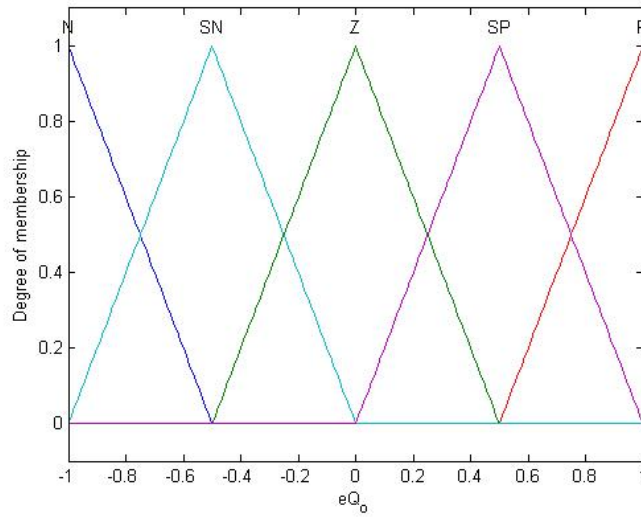


Figure 3.8 – Membership Functions For Input 1 Of Flow Control, eQ_o

Input 2, ceQ_o : For this input, three membership functions are used. They are labelled *Negative* (N), *Zero* (Z) and *Positive* (P). This is the rate of change in the error and gives an indication of how fast the difference in actual flow rate and the set point changes. The range for this input is $[-1 \ 1]$. The membership functions are triangular and defined as follows: N is $[-2 \ -1 \ 0]$, Z is $[-1 \ 0 \ 1]$ and P is $[0 \ 1 \ 2]$. The Membership Functions of Input 2 for Flow Control are given in Figure 3.9.

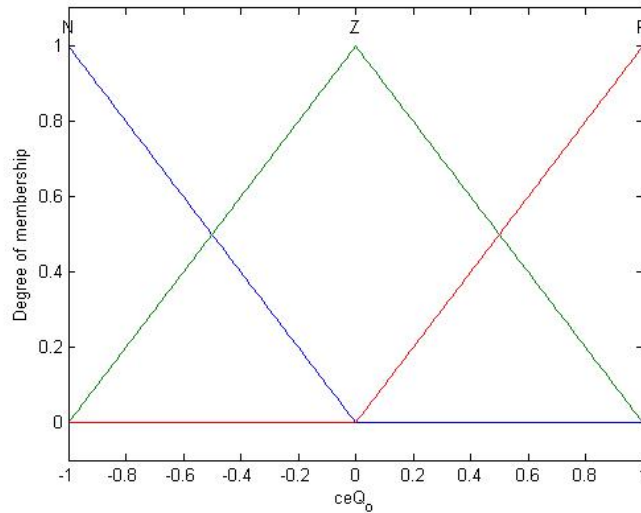


Figure 3.9 – Membership Functions For Input 2 Of Flow Control, ceQ_o

Output 1, Q_{in} : The output function is the input flow rate and is described by seven membership functions. They are labelled *Big Negative* (BN), *Negative* (N), *Small Negative* (SN), *Zero* (Z), *Small Positive* (SP), *Positive* (P) and *Big Positive* (BP). These

functions are triangular, with specifications: BN is $[-1.5 \ -1 \ -0.5]$, N is $[-1 \ -0.5 \ 0]$, SN is $[-0.5 \ -0.25 \ 0]$, Z is $[-0.05 \ 0 \ 0.05]$, SP is $[0 \ 0.25 \ 0.5]$, P is $[0 \ 0.5 \ 1]$ and BP is $[0.5 \ 1 \ 1.5]$. Figure 3.10 shows the Membership Functions of Output 1 for Flow Control.

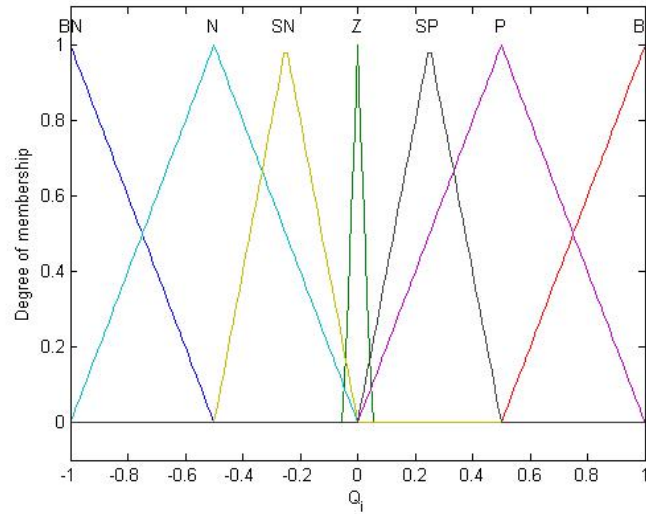


Figure 3.10 – Membership Functions For Output 1 Of Flow Control, Q_{in}

The reason for using more than three membership functions in Input 1 and the output is that faster, more defined control was needed. The rules were then adapted to accommodate the new sets. The rules used for flow control is shown in Table 3.1:

Error	N	SN	Z	SP	P
cError					
N	BN	BN	N	SP	Z
Z	N	N	Z	P	P
P	Z	SN	P	BP	BP

Table 3.1 – Fuzzy Logic Rules For Flow Control

3.4.2 Fuzzy Logic: Temperature Control

The inputs to the Temperature controller is chosen as the error in output temperature, eT_o and the change in error in output temperature, ceT_o . The change in power, dW , is the output to the controller. The membership functions used are:

Input 1, eT_o : Five membership functions are used to describe the error in output temperature, labelled *Negative Big* (NB), *Negative Small* (NS), *Zero* (Z), *Positive Small* (PS) and *Positive Big* (PB). The range is [-1 1]. Triangular membership functions are used. Their specifications are: NB is [-1.5 -1 -0.5], NS is [-1 -0.5 0], Z is [-0.5 0 0.5], PS is [0 0.5 1] and PB is [0.5 1 1.5]. Figure 3.11 shows the Membership Functions of Input 1 for Temperature Control.

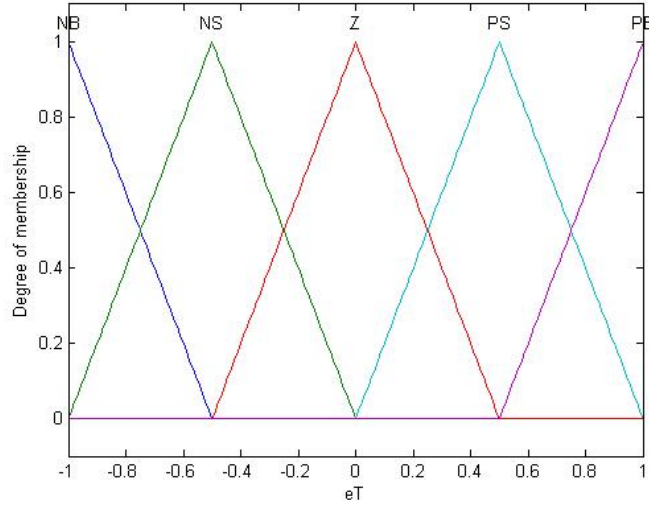


Figure 3.11 – Membership Functions For Input 1 Of Temperature Control, eT_o

Input 2, ceT_o : The change in output temperature error is described by three membership functions. They are labeled *Negative* (N), *Zero* (Z) and *Positive* (P). The range for this input is [-1 1]. The membership functions are triangular and defined as follows: N is [-2 -1 0], Z is [-1 0 1] and P is [0 1 2]. The Membership Functions of Input 2 for Temperature control are shown in Figure 3.12.

Output 1, cW : The change in power needed to obtain the correct temperature is described by five membership functions. They are labeled *Negative Big* (NB), *Negative Small* (NS), *Zero* (Z), *Positive Small* (PS) and *Positive Big* (PB). These triangular functions are specified as: NB is [-1.5 -1 -0.5], NS is [-1 -0.5 0], Z is [-0.1 0 0.1], PS is [0 0.5 1] and PB is [0.5 1 1.5]. These Membership Functions for Output 1 of Temperature Control are shown in Figure 3.13.

The Fuzzy Rules are presented in Table 3.2:

3.5 Model Reference Adaptive Control

In this paragraph, the design of the Model Reference Adaptive Controller will be described. This controller adapts its controller gains to suit the system. This is very useful in non-linear systems such as the one described in this problem. This means

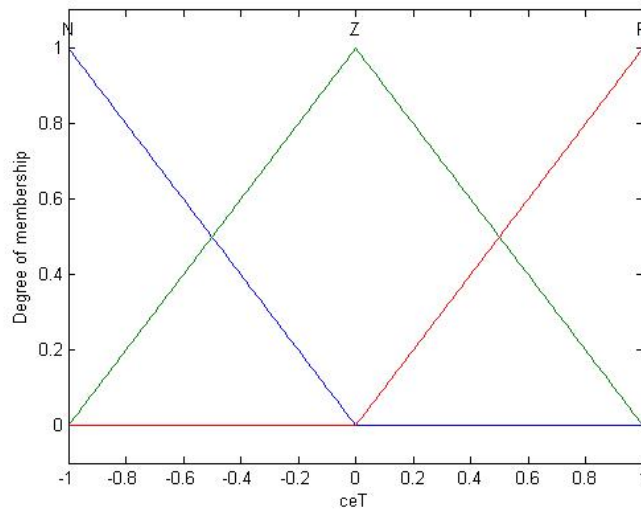


Figure 3.12 – Membership Functions For Input 2 Of Temperature Control, ceT_o

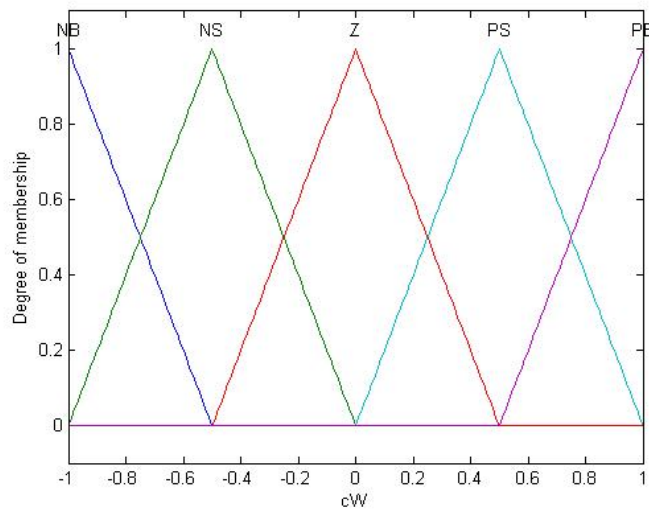
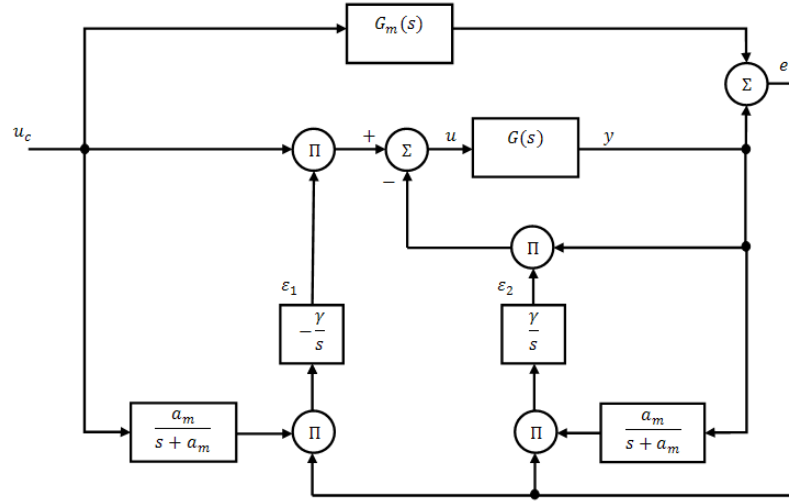


Figure 3.13 – Membership Functions For Output 1 Of Temperature Control, cW

that where a linearised controller will only work sufficiently at a certain work point, this adaptive controller will be able to control sufficiently at values different from the nominal case. In the situations where the heights of the levels of liquid in the tanks and the flow rates are at the working points, this controller will not necessarily improve the performance, but when the values of the heights and flow rates are at values different from the nominal values, the MRAC is expected to improve the controller performance. To determine the gain functions, the MIT rule is used. See section 2.3 on MRAC. Figure 3.14 shows the block diagram of the MRAC for the system.

To design an MRAC controller, the following should be chosen: A reference model, a controller structure and the tuning gains. A first order reference model is chosen and

Error	NB	NS	Z	PS	PB
cError					
N	NB	NS	NS	Z	PS
Z	NB	NS	Z	PS	PB
P	NS	Z	PS	PS	PB

Table 3.2 – Fuzzy Logic Rules For Temperature Control**Figure 3.14** – Block Diagram Of A Model Reference Adaptive Controller

the tuning gains are determined experimentally.

The transfer function of a first order reference model is defined as $G(s) = \frac{a_m}{s+a_m}$, which gives unity gain and has a time constant, $\tau = \frac{1}{a_m}$. The flow controller is designed for a 2% settling time of 40 sec, which means $4\tau = 2\%ts = 40$ sec, thus $\tau_{flow} = 10$ sec and $a_m = 0.1$ for flow control. For temperature control the desired 2%ts is 100 sec. Therefore $\tau_{temp} = 25$ sec and $a_m = \frac{1}{\tau_{temp}} = 0.04$ for temperature control.

The linearised temperature transfer function of the plant, is of the form $G(s) = \frac{y}{u} = \frac{b}{s+a}$, from equation 3.38 $b = 0.0009556$ and $a = 0.04$. These functions do not take the effect of the temperature of tank one into account. This effect is shown in equation 3.25. To compensate for this effect, the term, $l_3\theta_1$, is added to the control signal. The control signal is multiplied by $\frac{1}{C_2}$, therefore the added term is multiplied by C_2 . Figure 3.15 shows a diagram for the MRAC for temperature control with the term $l_3C_2\theta_1$ added to the control signal. The control equation is now $u = \epsilon_1 u_c - \epsilon_2 y + l_3 C_2 \theta_1$, therefore $W = \epsilon_1 \theta_{SP} - \epsilon_2 \theta_2 + l_3 C_2 \theta_1$. The transfer function of the model is $G_m(s) = \frac{y_m}{u_c} = \frac{b_m}{s+a_m} = \frac{0.04}{s+0.04}$. To determine ϵ_{temp1} and ϵ_{temp2} , the closed loop transfer function, $G_{CL}(s) = \frac{y}{u_c} = \frac{b\epsilon_{temp1}}{s+a+b\epsilon_{temp2}}$ is set equal to the transfer function of the model. When $G_{CL}(s) = G_m(s)$, $b\epsilon_{temp1} = b_m = 0.04$ and $a + b\epsilon_{temp2} = a_m = 0.04$. This yields $\epsilon_{temp1} = 41.86$ and $\epsilon_{temp2} = 0$. These initial values are at a flow rate of $0.01 \text{ m}^3/\text{sec}$ throughout the system.

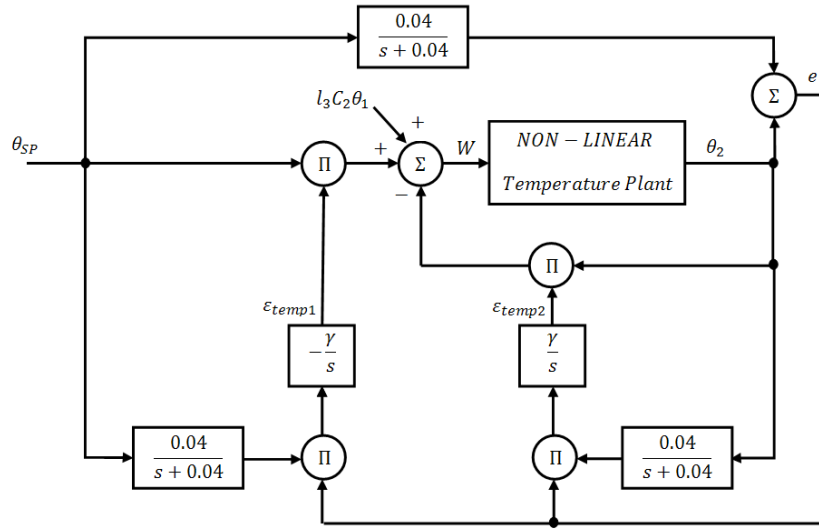


Figure 3.15 – Block Diagram Of The Model Reference Adaptive Controller For Temperature Control

The linearised open loop transfer function for the flow plant is a second order function with real poles at -0.4134 and -0.019 and unity gain (equations 3.28 and 3.29). The pole at -0.019 is the dominant pole and the second order function can be approximated by a first order function with unity gain and a pole at -0.019 . The transfer function of the model is $G_m(s) = \frac{y_m}{u_c} = \frac{b_m}{s+a_m} = \frac{0.1}{s+0.1}$. The closed loop transfer function of the plant is $G_{CL}(s) = \frac{y}{u_c} = \frac{b\epsilon_{flow1}}{s+a+b\epsilon_{flow2}}$, from the control equation, $u = \epsilon_1 u_c - \epsilon_2 y$. To determine the initial values for ϵ_{flow1} and ϵ_{flow2} , $G_{CL}(s) = G_m(s)$. The initial values of ϵ_{flow1} and ϵ_{flow2} for the flow control is 5.263 and 4.263 respectively.

The values of the tuning gains are $\gamma_{flow} = 0.5$ and $\gamma_{temp} = 0.001$. These values are obtained, by inserting pulse signals as set point values for both the flow and temperature controllers. At these chosen gains, the plant outputs follow the reference model outputs accurately and the values of ϵ_1 and ϵ_2 changed smoothly at flow rates different from the nominal value of $0.01 \text{ m}^3/\text{sec}$.

The results of the flow control are shown in Figure 3.16 at $0.01 \text{ m}^3/\text{sec}$. The values of ϵ_{flow1} and ϵ_{flow2} are constant at the initial values of 5.263 and 4.263, even if the flow rate changes. The reason for this is that a change in flow rate changes the flow equations marginally. At a flow rate of $0.015 \text{ m}^3/\text{sec}$, the linearised transfer function for the flow changed to $G(s) = \frac{0.0188}{s+0.0188}$ from $G(s) = \frac{0.019}{s+0.019}$. The values of ϵ_{flow1} and ϵ_{flow2} therefore do not change when the flow rates change. When the initial values for ϵ_{flow1} and ϵ_{flow2} are inserted as 5.26 and 4.26 respectively, the values change to the correct values of 5.263 and 4.263. This is shown in Figure 3.17. At a flow rate of $0.0075 \text{ m}^3/\text{sec}$, the outputs from the plant and the model are shown in 3.18.

The temperature controller is implemented and Figure 3.19 shows the output temperatures from the plant and the model at the nominal flow rates. At a flow rate of

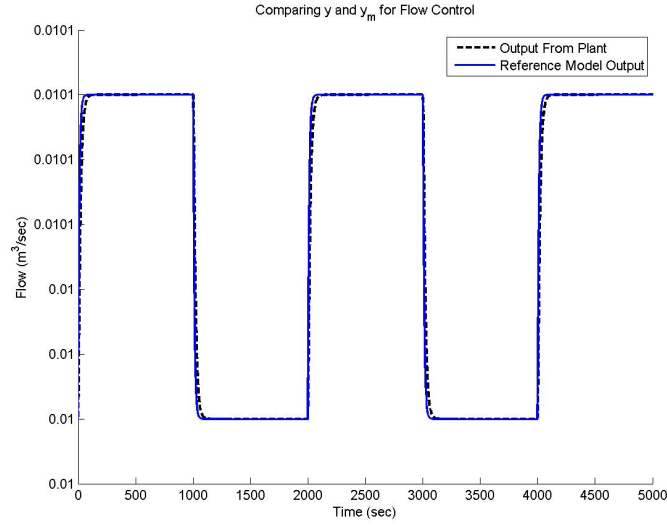


Figure 3.16 – MRAC For Flow Control With Pulse Command Signal, $Q = 0.01 \text{ m}^3/\text{sec}$, $\gamma_{flow} = 0.5$

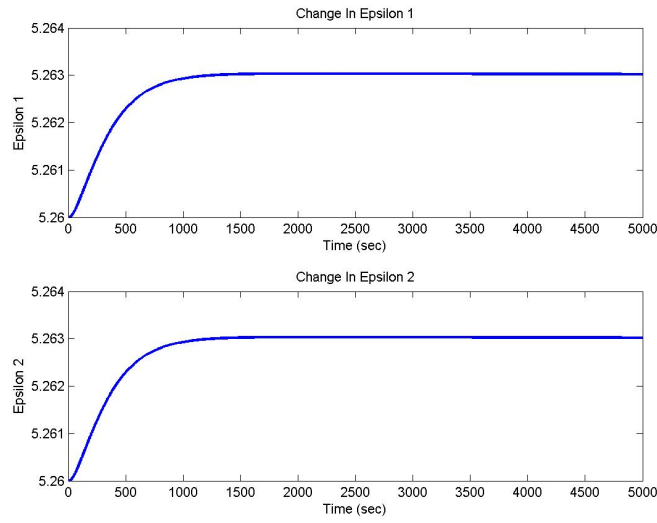


Figure 3.17 – Change In ϵ_{flow1} And ϵ_{flow2} With Pulse Command Signal, $Q = 0.01 \text{ m}^3/\text{sec}$, $\gamma_{flow} = 0.5$

$0.015 \text{ m}^3/\text{sec}$, the output temperature of the plant and the model is shown in Figure 3.21 and the change in ϵ_{temp1} and ϵ_{temp2} are shown in Figure 3.20. The control parameters, ϵ_{temp1} and ϵ_{temp2} changes to new values at different flow rates, as shown in Figure 3.20, but small errors between the model and the plant exist at values away from the nominal value. This error signal is shown in Figure 3.22. The error signal is large while the values of ϵ_{temp1} and ϵ_{temp2} changes, but after 500 sec, the error signal is very small and symmetric around 0.

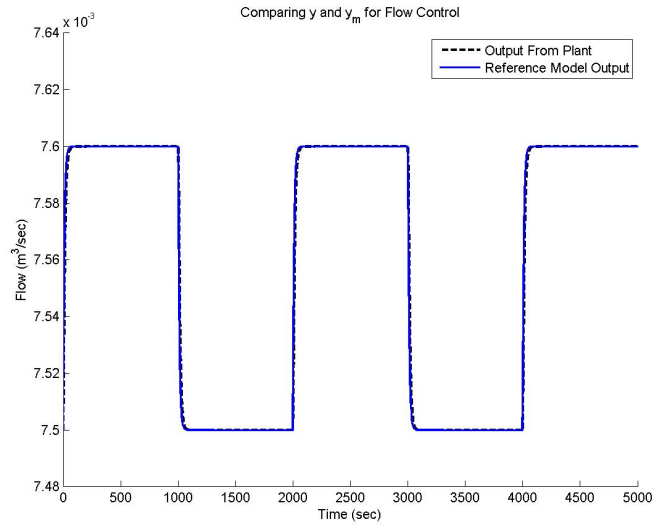


Figure 3.18 – MRAC For Flow Control With Pulse Command Signal, $Q = 0.0075 \text{ m}^3/\text{sec}$, $\gamma_{flow} = 0.5$

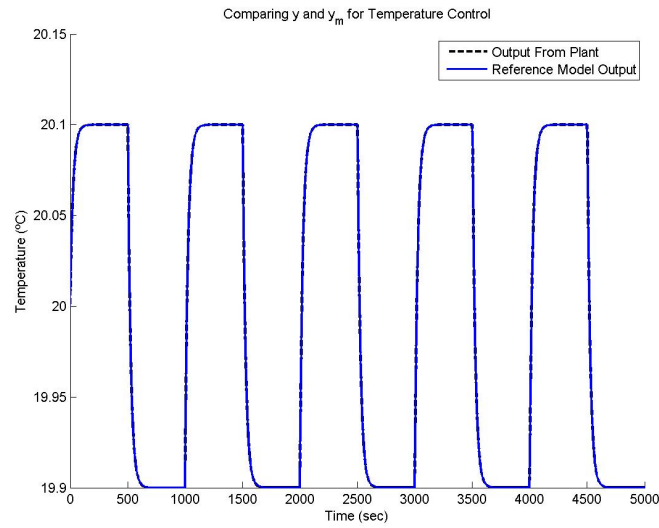


Figure 3.19 – MRAC For Temperature Control With Pulse Command Signal, $Q = 0.01 \text{ m}^3/\text{sec}$, $\gamma_{temp} = 0.001$

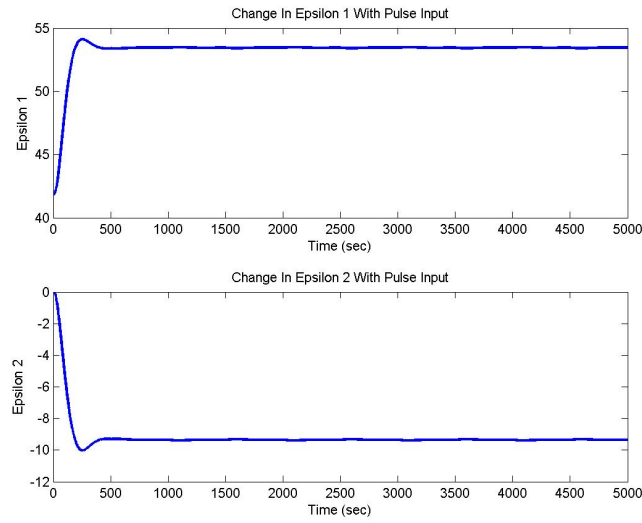


Figure 3.20 – Change In ϵ_{temp1} And ϵ_{temp2} With Pulse Command Signal, $Q = 0.015 \text{ m}^3/\text{sec}$, $\gamma_{temp} = 0.001$

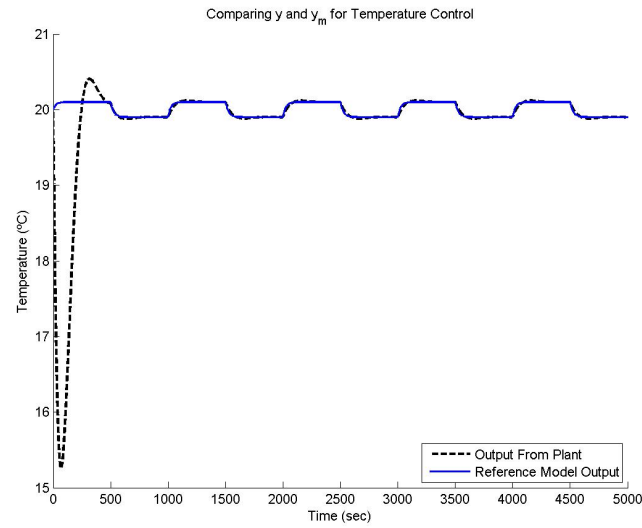


Figure 3.21 – MRAC For Temperature Control With Pulse Command Signal, $Q = 0.015 \text{ m}^3/\text{sec}$, $\gamma_{temp} = 0.001$

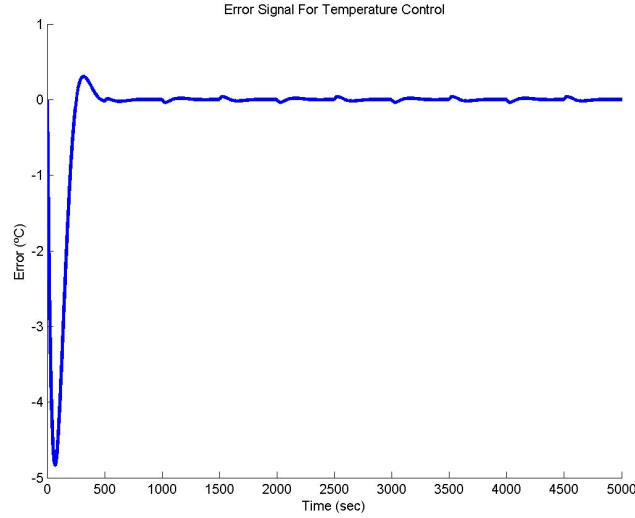


Figure 3.22 – Error Signal For Temperature MRAC, $Q = 0.015 \text{ m}^3/\text{sec}$

The reason for the design of the Model Reference Adaptive Controllers is to increase controller performance when the plant operates at flow rate values away from the nominal values at which it is linearised. In the case of the flow control, the difference in the model is too small to have an effect on the controller parameters. Therefore it is not expected that the MRAC for flow control will give better results than the Linearised or Fuzzy Logic controllers at flow rates away from the nominal work points. For temperature control, the controller parameters change when the plant operates at different flow rates, but an error occurs and the plant output deviates from the model output at flow rates away from the nominal work point. The further the flow rate is from the nominal flow rate, the bigger the difference between the plant output and the model output becomes.

3.6 Results And Comparison Between Controller Performances

In this paragraph the results of three different controllers are compared: The Linearised controller, Fuzzy Logic controller and the MRAC (Model Reference Adaptive Controller). Their responses to set point changes at different flow rates as well as disturbances in the input flow rate and input temperature are shown.

The initial values are chosen as follows: $\theta_{in} = 10^\circ\text{C}$, $\theta_1 = 10^\circ\text{C}$ and $\theta_2 = 20^\circ\text{C}$. The initial values of H_1 and H_2 are given for different flow rates in table 3.3.

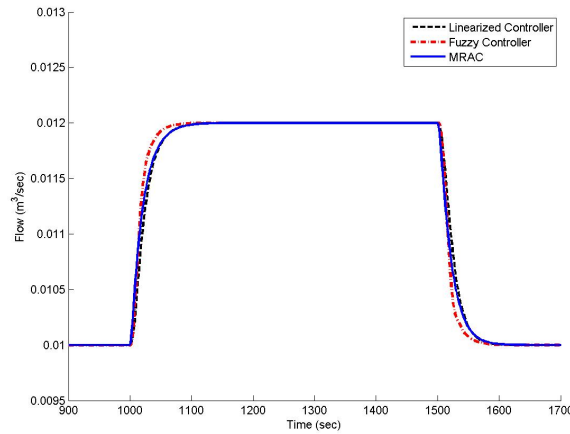
The different controllers are first compared without any disturbances in the input flow rate or temperatures (case 1). Set point changes in the flow rate or temperature occur respectively and the results are shown at different flow rates. In cases 2 and 3, disturbances occur in the input flow rate (case 2) and the input temperature (case 3).

Flow rate (m^3/sec)	H_1 (m)	H_2 (m)
0.01	0.301	0.25
0.015	0.639	0.5625
0.0075	0.1789	0.1406
0.005	0.088	0.0625

Table 3.3 – Initial Values Of H_1 and H_2 At Different Flow Rates

3.6.1 Case 1: No Disturbances

In this section, no disturbances occur in the input flow or temperature. The responses of the output flow rate and output temperature are shown when set point changes occur. The temperature set point is set to $\theta_{SP} = 30^\circ\text{C}$ at 2000 sec and is reset to 20°C at 2500 sec. The flow rate set point is increased with 20% at each different flow rate after 1000 sec, and changed to its original value after 1500 sec. Figures 3.23 to 3.25 show the flow responses of all three controllers with changes in flow rate set point values. The effect of a change in flow rate set point on the temperature is similar at different flow rates and is shown in Figure 3.26 at a flow rate of $0.01 \text{ m}^3/\text{sec}$. Figures 3.27 to 3.29 show the temperature responses with changes in temperature set point values at different flow rates.

**Figure 3.23** – Case 1: Output Flow At $Q = 0.01 \text{ m}^3/\text{sec}$

Set point changes in flow rate, Flow Response: The three different controllers perform similarly with a set point change in the flow rate, especially at the nominal flow rate of $0.01 \text{ m}^3/\text{sec}$, Figure 3.23. At a higher flow rate of $0.015 \text{ m}^3/\text{sec}$, the Fuzzy Logic controller has a smaller settling time than the Linearised controller or the MRAC, as shown in Figure 3.24. At a lower flow rate of $0.005 \text{ m}^3/\text{sec}$, the MRAC is the only controller of which the response does not change at this low flow rate. The Linearised controller now shows a small overshoot. This is shown in Figure 3.25. Despite these changes, all the controllers can follow the set point changes in the flow rate very well at different flow rates.

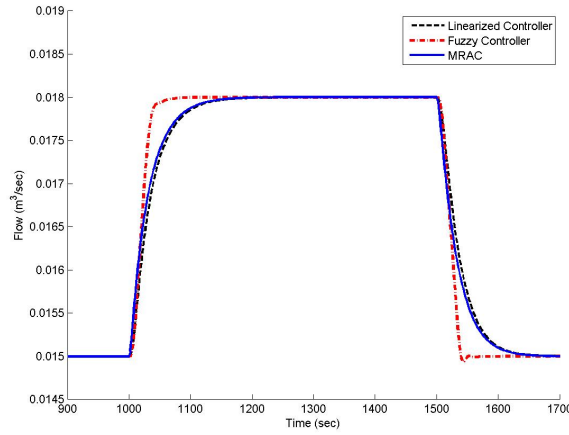


Figure 3.24 – Case 1: Output Flow At $Q = 0.015 \text{ m}^3/\text{sec}$

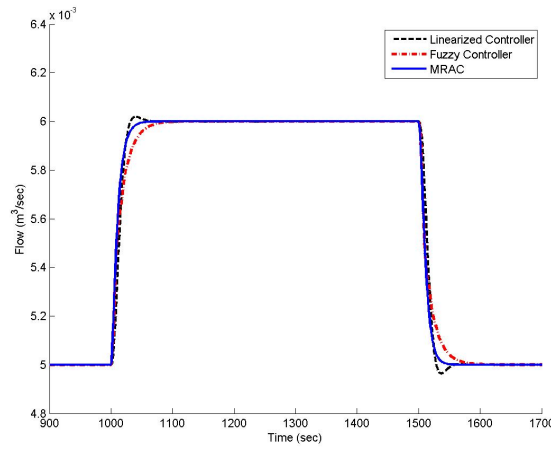


Figure 3.25 – Case 1: Output Flow At $Q = 0.005 \text{ m}^3/\text{sec}$

Set point changes in flow rate, Temperature Response: The effect of a change in flow rate set point on the temperature is shown in Figure 3.26. The MRAC shows the biggest effect from the change in flow rate set point with a maximum deviation of 4.4°C . The Linearised controller shows the smallest deviation from its value with a change in flow rate set point. Both the Fuzzy Logic controller and the Linearised controller have smaller deviations and recover within 200 sec, while the MRAC takes 300 sec to recover.

Set point changes in temperature, Temperature Response: All three temperature controllers show sufficient results for a change in temperature set points. The set point changes are done at the nominal flow rate value, $0.01 \text{ m}^3/\text{sec}$, at a flow rate higher than the nominal value, $0.015 \text{ m}^3/\text{sec}$, and at a flow rate lower than the nominal value, $0.0075 \text{ m}^3/\text{sec}$. At a flow rate of $0.01 \text{ m}^3/\text{sec}$ the results from Figure 3.27 shows that the three different controllers all have a similar $2\%t_s$ of 130 sec. The MRAC has no overshoot. The maximum overshoot from the Linearised controller and the Fuzzy Logic controller, is 0.41°C and 0.25°C respectively. At a flow rate of $0.015 \text{ m}^3/\text{sec}$, the

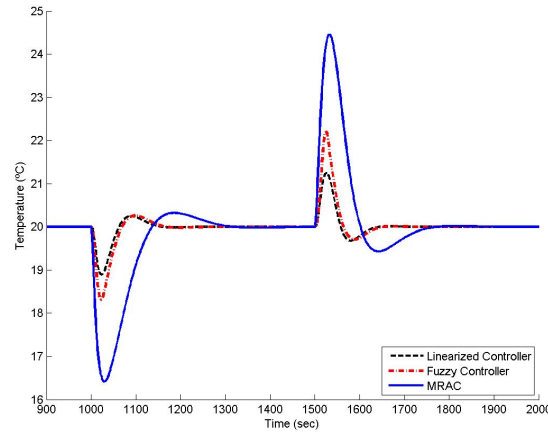


Figure 3.26 – Case 1: Effect Of Flow Rate Change On Temperature At $Q = 0.01 \text{ m}^3/\text{sec}$

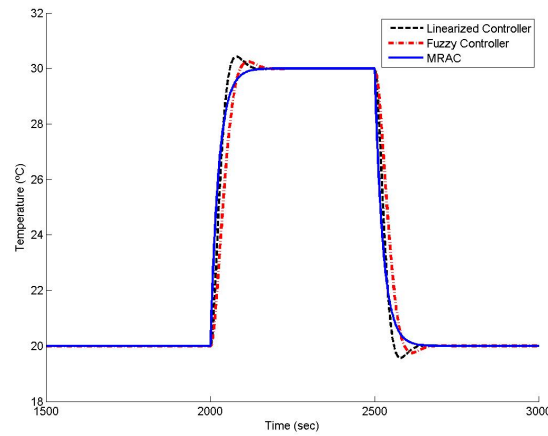


Figure 3.27 – Case 1: Output Temperature At $Q = 0.01 \text{ m}^3/\text{sec}$

maximum overshoots and the settling times of all three controllers increased. The Linearised controller now shows the shortest 2% settling time of 157 sec. The MRAC settles the slowest, with $2\%t_s = 270 \text{ sec}$. This is shown in Figure 3.28. Figure 3.29 shows the results at a flow rate of $0.0075 \text{ m}^3/\text{sec}$. In this case the Linearised controller settles with $2\%t_s = 53 \text{ sec}$. For the Fuzzy Logic controller, $2\%t_s = 77 \text{ sec}$ and for the MRAC, $2\%t_s = 118 \text{ sec}$. None of the controllers shows a big overshoot.

3.6.2 Case 2: Disturbances In Q_{in}

The effect of a pulse disturbance in the input flow rate is evaluated. A pulse disturbance of 10% of the flow rate is inserted after 2000 sec. The duration of the pulse is 500 sec.

The effect of a disturbance in input flow rate on both the flow rate and temperature is similar at different flow rates and is shown at $Q = 0.01 \text{ m}^3/\text{sec}$ in Figures 3.30 and 3.31.

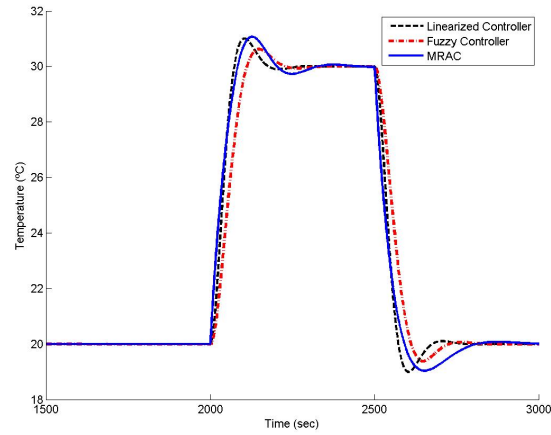


Figure 3.28 – Case 1: Output Temperature At $Q = 0.015 \text{ m}^3/\text{sec}$

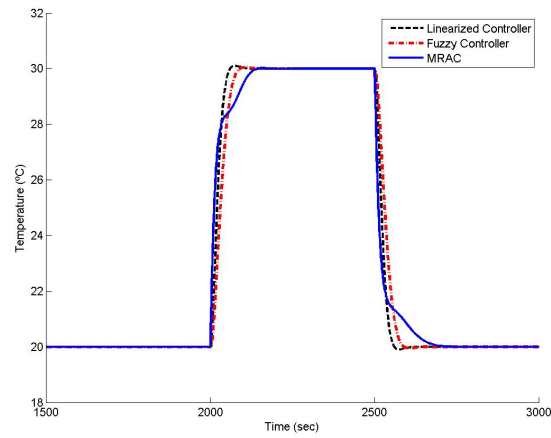


Figure 3.29 – Case 1: Output Temperature At $Q = 0.0075 \text{ m}^3/\text{sec}$

Effect of input flow disturbances on output flow rate: In Figure 3.30, it is clear that the MRAC shows the biggest effect from a disturbance in the input flow. The output flow deviates from the set point value with $0.00017 \text{ m}^3/\text{sec}$. With the Fuzzy Logic controller, the smallest deviation occurs and has a value of $0.00002 \text{ m}^3/\text{sec}$. The Linearised controller shows a deviation of $0.00003 \text{ m}^3/\text{sec}$. For short disturbances, these effects are very small, but the longer the duration of the disturbance, the worse the effect with the MRAC will be.

Effect of input flow disturbances on output temperature: The effect of a flow disturbance on the output temperature is shown in Figure 3.31. The MRAC shows the biggest effect with a deviation of 0.38°C and takes 300sec to recover from the disturbance. The effect on the Linearised controller is much smaller with a maximum deviation of 0.05°C . The Fuzzy Logic shows the smallest effect with a deviation of 0.03°C . The Linearised controller and the Fuzzy Logic controller recover within 150sec from the disturbance.

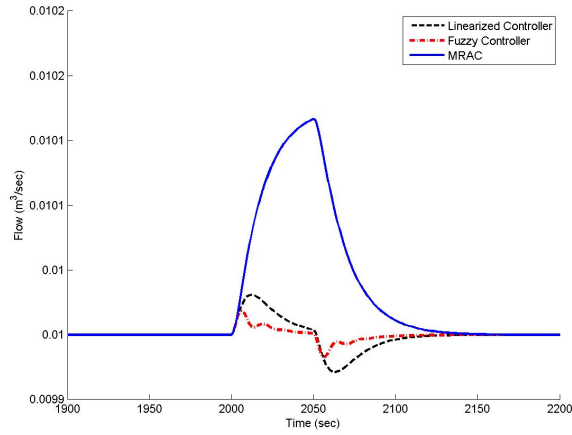


Figure 3.30 – Case 2: Output Flow At $Q = 0.01 \text{ m}^3/\text{sec}$

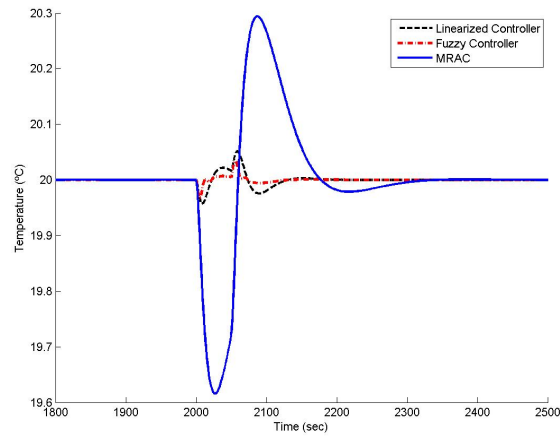


Figure 3.31 – Case 2: Output Temperature at $Q = 0.01 \text{ m}^3/\text{sec}$

3.6.3 Case 3: Disturbances In θ_{in}

The third comparison shows the effect of a pulse disturbance in the input temperature. After 2000 sec, the input temperature increases with 10%, and decreases again 500 sec later at 2500 sec. The set point value for output temperature stays at 20 °C. The effect on the output temperature is similar at different work points and is shown in Figure 3.32 at a flow rate of $0.01 \text{ m}^3/\text{sec}$.

Effect of input temperature disturbances on output temperature: Figure 3.32 shows that a disturbance in the input temperature has no effect on the output temperature when an MRAC is used. The reason is that the MRAC takes the temperature from tank one into account when determining the control signal. The deviation from 20 °C with the Linearised controller is $0.19 \text{ }^\circ\text{C}$ and the Fuzzy Logic controller has a deviation of $0.26 \text{ }^\circ\text{C}$.

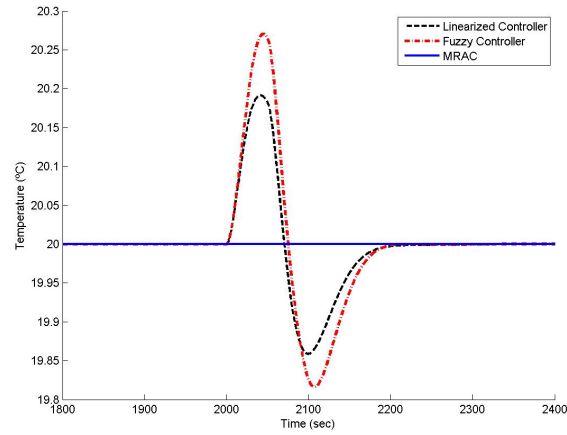


Figure 3.32 – Case 3: Output Temperature At $Q = 0.01 \text{ m}^3/\text{sec}$

3.7 Conclusion

The model of the tank that was created in this section simulated the flow of water through the tank correctly and the tank model can be used in further simulations.

The results show that all three controllers can control the output flow and temperature to certain set point values. When disturbances are added, the Fuzzy Controller shows the best disturbance rejection for disturbances in input flow rate. For temperature control, the disturbance has no effect when using the MRAC. In this case a disturbance in input temperature is rejected perfectly with the MRAC. With disturbances or set point changes in flow rate the MRACs for temperature and flow control reject disturbances poorly. If flow disturbances occur for a much longer time than the 500 sec pulses, the MRAC will not be sufficient for flow disturbance rejection.

The different control methods used on the simplified model, provided the necessary knowledge and these control methods can now be implemented on a more realistic representation of the cascaded tank model.

Chapter 4

Height Control Of Buffers

In this chapter, the height control of the system is discussed. This is done to control the buffer. In the original factory, no buffering, except for the pipelines between sub-plants, exists. The results from controlling the buffer will show the effect of disturbances in the output when the buffer capacities are not used efficiently to reject the disturbances. In the representation of the cascaded system, the buffers between different processes are modeled as tanks. This representation is shown in Figure 4.1.

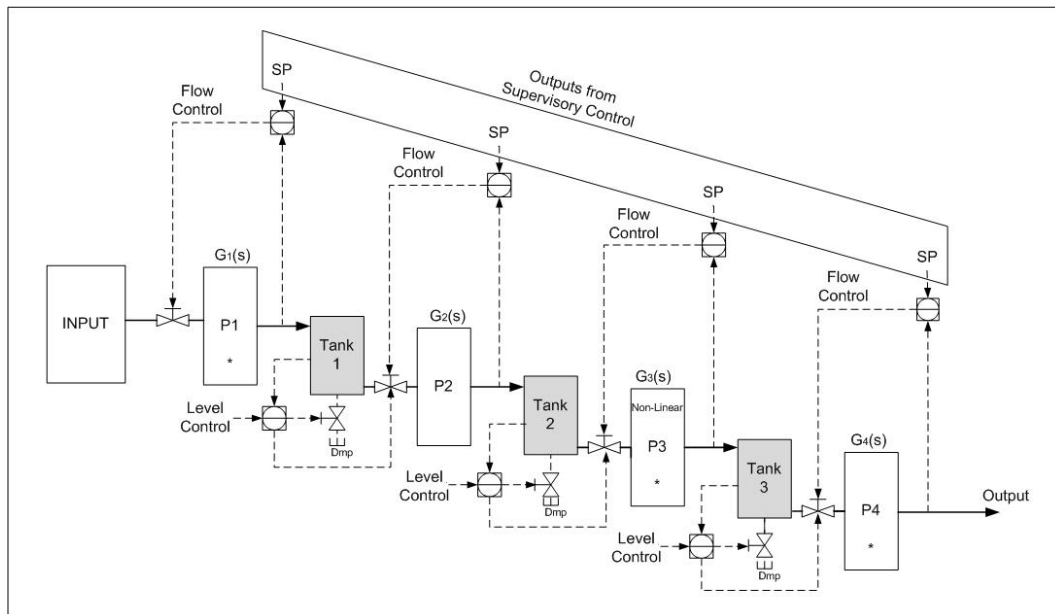


Figure 4.1 – Multiple Tank System

4.1 Controller Design

4.1.1 Parameters

Tank parameters: The base area of the tanks is chosen as 2 m^2 and the maximum height is 2 m . 100% flow input is $0.02\text{ m}^3/\text{sec}$. Valve position (VP) vary from 0 to 2. The nominal values for the height (H_{nom}), as well as for the flow (Q_{nom}), is 50%. This means the tanks have a nominal height of $H_{nom} = 1\text{ m}$ and a nominal flow $Q_{nom} = 0.01\text{ m}^3/\text{sec}$ throughout the whole plant system. To achieve these nominal values, the valves have a nominal value of $VP = 50\% = 1$. The linearised transfer function of the tank model, when looking at the change in height as a result of a change in valve position, is as follows:

$$G_{tank}(s) = \frac{H(s)}{VP(s)} = \frac{B}{\tau s + 1} = \frac{-2}{400s + 1} \quad (4.1)$$

The negative sign of the transfer function indicates that the height will increase if the value of the valve position decreases. This transfer function's parameters, as a result of linearisation, are at a certain work point. This work point was chosen at the nominal values of $Q = Q_{nom} = 50\%$ and $H = H_{nom} = 50\%$. To identify the model the valve position is increased by 1% with a step input, shown in Figure 4.2. It is then decreased by 1% with a step input and shown in Figure 4.3. In both cases 63, 2% of the final value is reached at 900sec, which gives a time constant of $\tau = 900 - 500 = 400\text{ sec}$. With an increase in the valve position the gain is $\frac{\Delta H}{\Delta VP} = -2.06$. With a decrease in valve position the gain is $\frac{\Delta H}{\Delta VP} = -1.94$. The bigger the step in valve position, the more the gain differs from -2 , but the mean between the step-up and step-down gains is -2 . Therefore the transfer function has a gain of -2 and a time constant of 400 sec at the chosen work points.

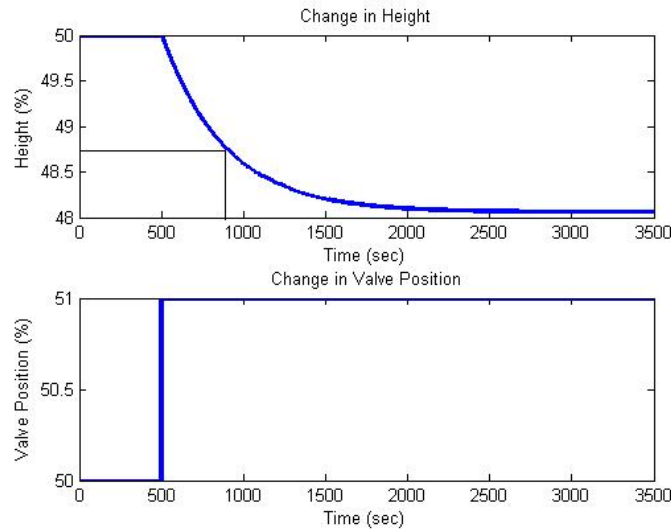


Figure 4.2 – Change In Height As A Result Of A Step Increase In Valve Position

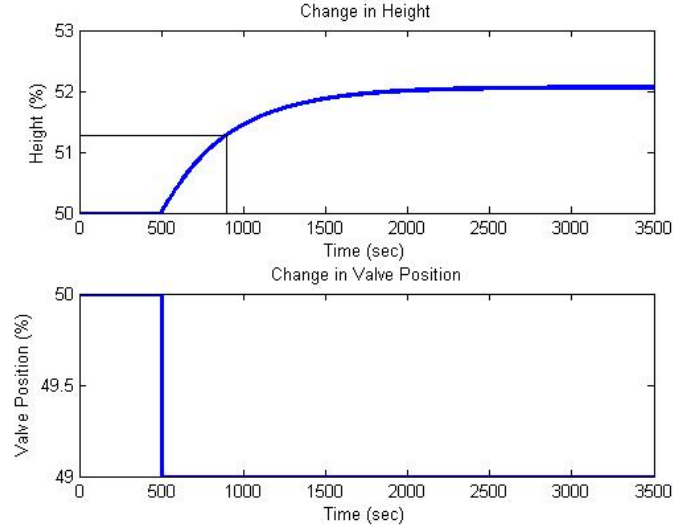


Figure 4.3 – Change In Height As A Result Of A Step Decrease In Valve Position

Plant Parameters: The system consists of four sub-plants with three tanks in-between them, see Figure 4.1.

These sub-plants can be described by the following transfer functions:

$$G_{plant1}(s) = \frac{Q_{out}}{Q_{in}} = \frac{k_1}{\tau_{p1}s + 1} e^{-\phi_{p1}s} \quad (4.2)$$

$$G_{plant2}(s) = \frac{Q_{out}}{Q_{in}} = \frac{k_2}{\tau_{p2}s + 1} e^{-\phi_{p2}s} \quad (4.3)$$

$$G_{plant3}(s) = \frac{Q_{out}}{Q_{in}} = \frac{k_3}{\tau_{p3}s + 1} e^{-\phi_{p3}s} \quad (4.4)$$

$$G_{plant4}(s) = \frac{Q_{out}}{Q_{in}} = \frac{k_4}{\tau_{p4}s + 1} e^{-\phi_{p4}s} \quad (4.5)$$

with

$$k_1 = k_2 = k_4 = 1$$

$$k_3 = \begin{cases} 75Q_{in} + 0.25 & \forall Q_{in} < Q_{nom} \\ 1 & \forall Q_{in} \geq Q_{nom} \end{cases} \quad (4.6)$$

where Q_{in} is the flow into sub-plant three. The gain of the third sub-plant is thus non-linear and determined by Q_{in} . When Q_{in} is lower than the nominal flow, Q_{nom} , the flow out of sub-plant three, Q_{out} , is less than Q_{in} .

The other plant parameters are:

$$\begin{aligned}
 \tau_{p1} &= 40 \text{ sec} \\
 \tau_{p2} &= 25 \text{ sec} \\
 \tau_{p3} &= 100 \text{ sec} \\
 \tau_{p4} &= 50 \text{ sec} \\
 \phi_{p1} &= 10 \text{ sec} \\
 \phi_{p2} &= 25 \text{ sec} \\
 \phi_{p3} &= 5 \text{ sec} \\
 \phi_{p4} &= 10 \text{ sec}
 \end{aligned}$$

4.1.2 Height Controllers

The height of tank one is not controlled. Tank one is used to absorb rapid changes in the input flow rate. The output flow from the first tank is set to follow the low pass filtered value of the input flow rate. A low pass filter with a time constant $\tau_{LP} = 400 \text{ sec}$, which is equal to the time constant of the tank model (equation 4.1), is used. The filtered flow rate is $Q_{LP}(s) = \frac{1}{400s+1} Q_{input}(s)$. This is done to make sure that the average flow rates in and out of each tank are the same, to prevent the tanks from overflowing or running empty and to absorb the higher frequency disturbances. The height of the first tank will vary to keep the outflow flow rate of tank one at the filtered input value. When the height of this tank exceeds the height limits for the tanks, changes are made to the output valves, to correct it. For a height above 80%, a dumping valve (VP_{dmp}) is activated, to act for the short time until the valve position, and therefore also the height, is changed. VP_{dmp} is deactivated (set to zero) when the height is below 75%. When the height goes below 20%, a cutback is activated, to decrease the valve opening and increase the height. The cutback is deactivated when the height is above 25%.

The objective is to keep the heights of all the tanks close enough to their nominal values of 50%, while the flow throughout the whole system follows the low pass filtered value of the input flow, Q_{LP} . Therefore height control is implemented on tanks two and three. The heights of tanks two and three are controlled with a PI (Proportional and Integral) controller in the base case and then with a Fuzzy Logic and a Model Reference Adaptive Controller. The heights are controlled to stay at a set point value, nominally at $H_{SP} = H_{nom} = 50\%$, despite changes in flow rate.

4.2 Base Case PI Control

To compare different controllers, base case control is first implemented. PI control is used for the base case control. PI control is sufficient for slow systems. The system can be noisy due to the equipment such as valves or sensors. The derivative term in this case can increase the noise. The PI control for height control is found fast enough, without oscillation. The other control methods, Fuzzy Logic control and Model Reference Adaptive control, are compared to the base case control in section 4.5. The transfer function for a PI controller is [4]:

$$G_c(s) = K(1 + \frac{1}{T_i s}) \quad (4.7)$$

A PI controller for the height control of tanks two and three is then designed to satisfy specific requirements. These requirements are the damping ratio, ζ , and a specific 2% settling time, $2\%t_s$. The damping ratio is chosen such that $\zeta = 0.707$. This damping ratio gives a relatively small overshoot, but a fast settling time. The dominant poles of the closed loop system are then a complex conjugate pair. The natural frequency is given by $\omega_n = \frac{4}{\zeta(2\%t_s)} = 0.0141$ rad/sec if the 2% settling time is chosen to be 400 sec.

The transfer function of each tank is modelled as a first order function of the form

$$G_{tank}(s) = \frac{B}{\tau s + 1} \quad (4.8)$$

where the gain is $B = -2$ around the working points of 50% flow and 50% height and the time constant is $\tau = 400$ sec. It can also be written as:

$$G_{tank}(s) = \frac{B}{\tau s + 1} = \frac{B/\tau}{s + 1/\tau} = \frac{b}{s + a} \quad (4.9)$$

where $b = B/\tau = -0.005$ and $a = 1/\tau = 0.0025$

The transfer function of the PI controller is of the form

$$G_c(s) = \frac{K_p s + K_i}{s} \quad (4.10)$$

with K_p and K_i the proportional and integral gains.

The closed loop transfer function is given by

$$T(s) = \frac{G_c(s)G_{tank}(s)}{1 + G_c(s)G_{tank}(s)} \quad (4.11)$$

And when equations 4.9 and 4.10 are substituted in 4.11 :

$$T(s) = \frac{b(K_p s + K_i)}{s(s + a) + b(K_p s + K_i)} = \frac{bK_p(s + K_i/K_p)}{s^2 + (a + bK_p)s + bK_i} \quad (4.12)$$

The denominator of $T(s)$ is equal to $s^2 + 2\zeta\omega_n s + \omega_n^2$, thus:

$$a + bK_p = 2\zeta\omega_n \quad (4.13)$$

$$\Rightarrow K_p = \frac{2\zeta\omega_n - a}{b} = -3.99 \quad (4.14)$$

$$bK_i = \omega_n^2 \quad (4.15)$$

$$\Rightarrow K_i = \frac{\omega_n^2}{b} = -3.976 \times 10^{-2} \quad (4.16)$$

These parameters were implemented and the controller is compared to other controllers. The results are shown in section 4.5.

4.3 Fuzzy Logic Control

This section describes the design of a fuzzy logic controller to control the heights of the levels of the second and third tanks. The height of the water level in the tank (H), the error in height ($eH = H - H_{SP}$) and the change in error in height (ceH) are the inputs to the controller and the output is the change in the valve position from the tank ($cValve$). Fuzzy Logic control was at first done without the derivative term, (ceH). The controller had a very slow response with a great overshoot. Oscillation occurred when the settling time was decreased. The addition of the derivative term, gave the controller response a fast settling time with a very small overshoot. The derivative term can however increase noise in the system.

4.3.1 Membership Functions

Each input and output consists of various membership functions and will be considered individually:

Input 1, H : The height of the water level in the tank can be either *Empty*, *Correct* or *Full*. These membership functions are shown in Figure 4.4

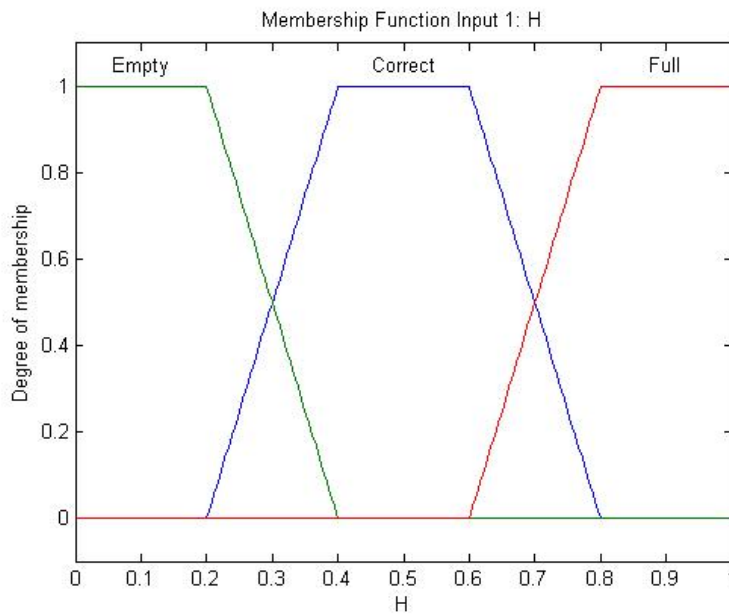


Figure 4.4 – Membership Functions Of Input 1: Height

Input 2, eH : The membership functions of eH are shown in figure 4.5. This shows that the error in the height can be *Negative*, *Zero* or *Positive*.

Input 3, ceH : To decrease settling time and overshoot, the input, ceH , is added. The membership functions of ceH are *Negative*, *Zero* or *Positive* and are shown in figure 4.6.

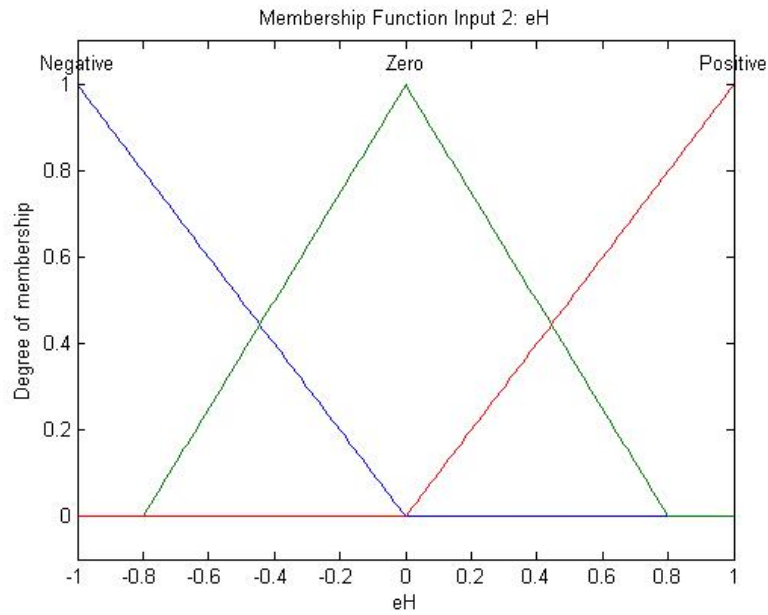


Figure 4.5 – Membership Functions Of Input 2: eH

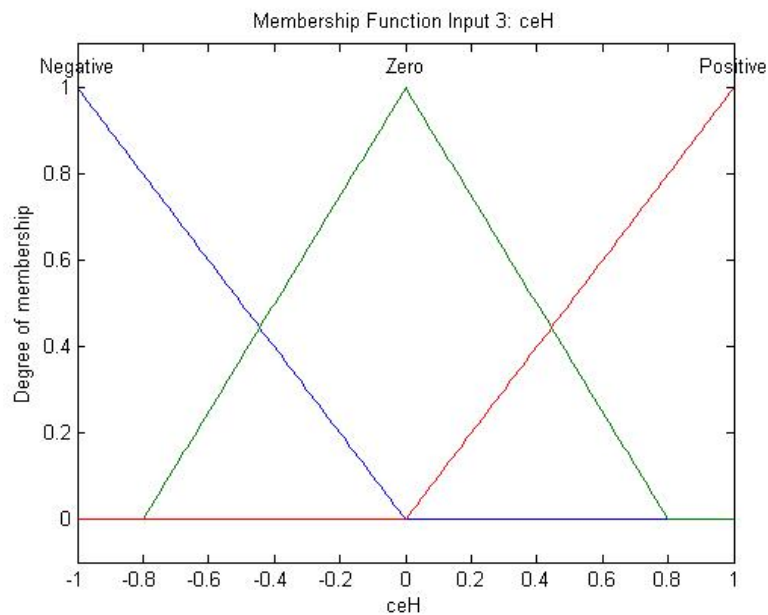


Figure 4.6 – Membership Functions Of Input 3: ceH

Output 1, $cValve$: The output from the fuzzy controller is the change in the position of the valve which controls the outflow from the tank. This change can be *Big Negative*, *Negative*, *Zero*, *Positive* and *Big Positive*. Figure 4.7 shows the membership functions of the change in valve position, $cValve$.

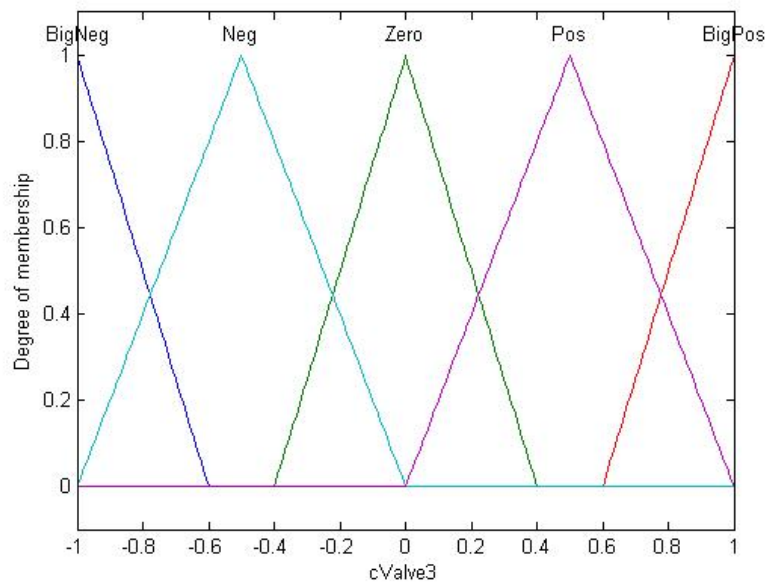


Figure 4.7 – Membership Functions Of Output 1: cValve

The inputs and output of the Fuzzy Logic Controller are Fuzzy variables. These variables are not in the same range as the actual inputs and outputs. To convert the actual inputs to fuzzy values, fuzzification is necessary and to convert the fuzzy output to the actual valve position, defuzzification is needed. The different fuzzification gains used for H , eH and ceH are 0.01, 0.05 and 1.5 respectively. The height is a percentage and is converted to [0 1]. The error in height is converted such that a 20% difference or more is a very big deviation (Fuzzy value -1 or 1). The defuzzification gain is equal to 0.2 for the rate of change in valve position.

4.3.2 Rules Of The Fuzzy Logic Controller

Table 4.1 shows the rules for the Fuzzy Logic Controller. These rules take both the height and the rate of change in height into account when determining the valve position.

The Fuzzy Logic controller is implemented and compared with the other controllers in section 4.5.

Height	eH	ceH	cValve
Empty	Negative	Negative	Big Negative
		Zero	Big Negative
		Positive	Negative
	Zero	Negative	Big Negative
		Zero	Negative
		Positive	Zero
	Positive	Negative	Negative
		Zero	Zero
		Positive	Positive
Correct	Negative	Negative	Big Negative
		Zero	Negative
		Positive	Zero
	Zero	Negative	Negative
		Zero	Zero
		Positive	Positive
	Positive	Negative	Zero
		Zero	Positive
		Positive	Big Positive
Full	Negative	Negative	Negative
		Zero	Zero
		Positive	Positive
	Zero	Negative	Zero
		Zero	Positive
		Positive	Big Positive
	Positive	Negative	Positive
		Zero	Big Positive
		Positive	Big Positive

Table 4.1 – Fuzzy Logic Rules For Height Control

4.4 Model Reference Adaptive Control

This section describes the third type of height control, namely Model Reference Adaptive Control. Figure 4.8 shows a block diagram of a MRAC to control the height of a buffer. To determine the gain functions, the MIT rule is used. It is described in section 2.3 and is also used in section 3.5.

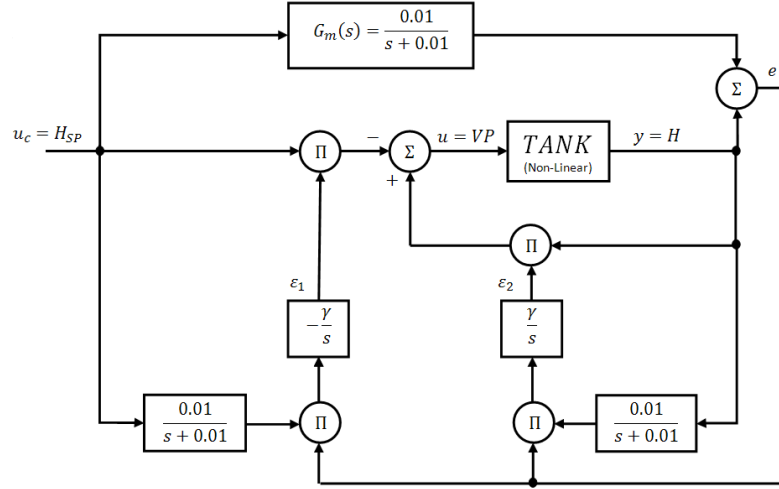


Figure 4.8 – Block Diagram Of The Model Reference Adaptive Controller For Height Control

To design an MRAC, a reference model, a controller structure and the tuning gains should be chosen. The chosen model to be followed can be written in the form $\frac{a_m}{s + a_m}$. This is a first order model with a settling time of 400 sec. The time constant of this system is given by $4 \times \tau_m = 2\%t_s = 400$ sec, with τ_m the time constant of the model and $2\%t_s$ the settling time of the model. Thus $\tau_m = 100$ sec $= \frac{1}{a_m}$. This yields $a_m = 0.01$.

To calculate initial conditions for ϵ_1 and ϵ_2 , the control equation from the diagram in Figure 4.8 as well as the linearised transfer function of the tank, from equation 4.9 is used. Figure 4.9 shows a simplified representation of Figure 4.8, to determine initial values for ϵ_1 and ϵ_2 .

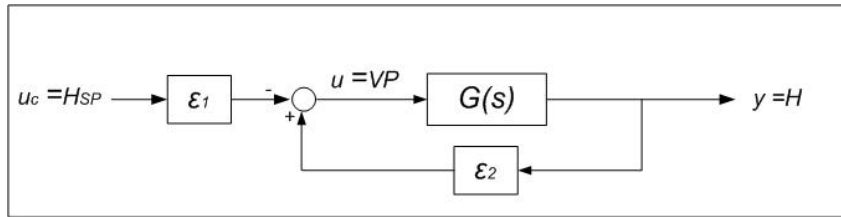


Figure 4.9 – Model Reference Adaptive Control Diagram To Calculate The Initial Values For ϵ_1 and ϵ_2

From equation 4.9:

$$y = \frac{-0.005}{s + 0.0025}u \quad (4.17)$$

From Figure 4.9:

$$u = -\epsilon_1 u_c + \epsilon_2 y \quad (4.18)$$

Therefore:

$$\frac{y}{u_c} = \frac{0.005\epsilon_1}{s + 0.0025 + 0.005\epsilon_2} \quad (4.19)$$

At the nominal values of $H = 50\%$ and $VP = 50\%$, equation 4.18 is:

$$\begin{aligned} VP &= -\epsilon_1 H_{SP} + \epsilon_2 H \\ \epsilon_2 &= \epsilon_1 + 1 \end{aligned} \quad (4.20)$$

With the gain from $\frac{y}{u_c}$ equal to the gain from $\frac{y_m}{u_c}$, the initial value for $\epsilon_1 = 2$. From equation 4.20, $\epsilon_2 = \epsilon_1 + 1 = 3$.

To determine γ , the rate at which the values of ϵ_1 and ϵ_2 adapts to the correct values is looked at. The initial values are $\epsilon_1 = 2$ and $\epsilon_2 = 3$. At a height other than the nominal 50%, these values of ϵ_1 and ϵ_2 should change. These values should also change when the flow rate through the system is different from the nominal flow rate of 50%.

To test the controller parameters, a pulse height signal with a 50% bias (set point of controlled variable, H) is entered as the controller input, u_c . The signal has an amplitude of 2% and a period of 2000 sec. With no disturbances in the plant input, Q_{in} , and a flow rate of 50%, the controller is tested. The results are shown in Figures 4.10 and 4.11.

For the next test, the set point is a pulse height signal around 75% with a flow rate of 50%. These results are shown in Figures 4.12 and 4.13. After this, a pulse signal is inserted as the set point around 50% height, but at a flow rate of 60%. Figures 4.14 and 4.15 show these results. In the last two cases, the rate at which the values of ϵ_1 and ϵ_2 change, are evaluated. For ϵ_1 and ϵ_2 to settle within 500 sec, without changing too fast to add an overshoot to the output, $\gamma = 1 \times 10^{-4}$.

The MRAC is implemented and the results are compared with the results of the other controllers in section 4.5.

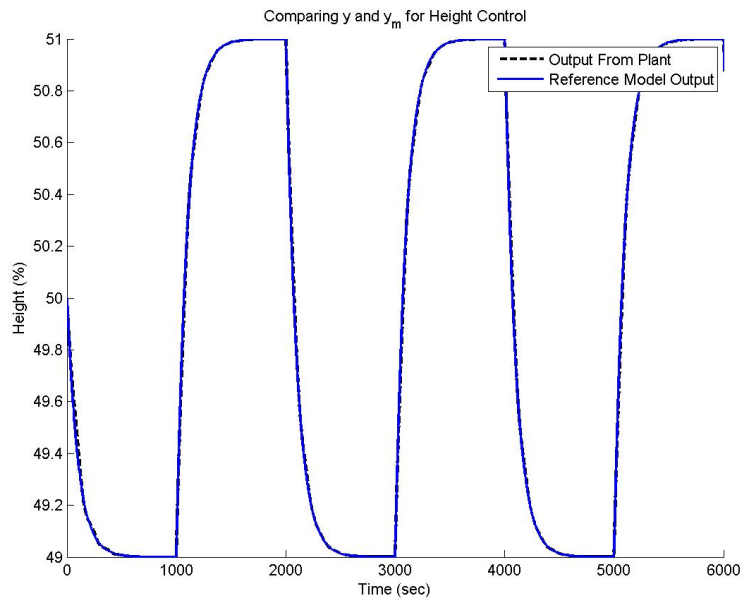


Figure 4.10 – Comparing Outputs From The Plant And The Model Of MRAC With Pulse Command Signal Around $H = 50\%$ With $Q = 50\%$

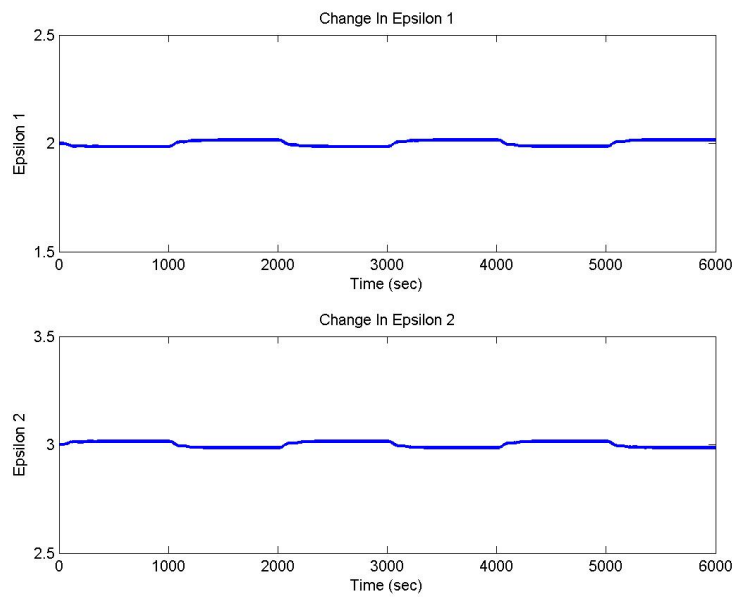


Figure 4.11 – Change in ϵ_1 And ϵ_2 As A Result Of Pulse Command Signal Around $H = 50\%$ With $Q = 50\%$

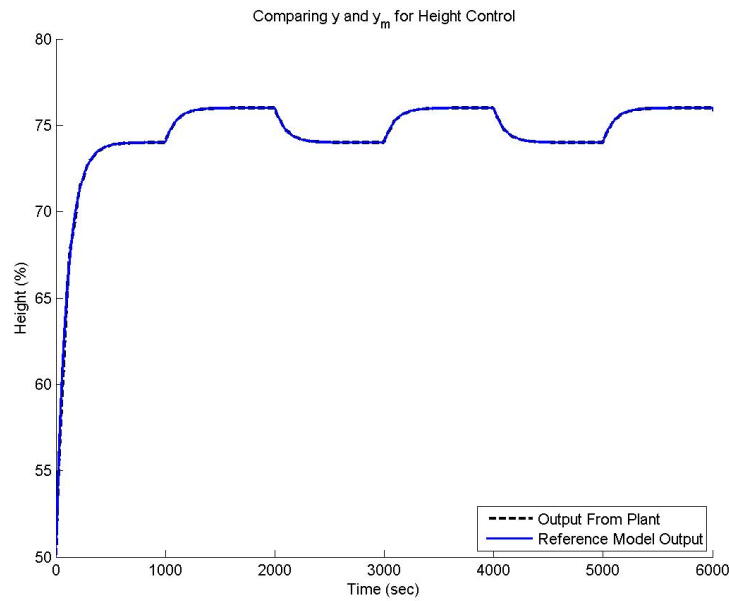


Figure 4.12 – Comparing Outputs From The Plant And The Model Of MRAC With Pulse Command Signal Around $H = 75\%$ With $Q = 50\%$

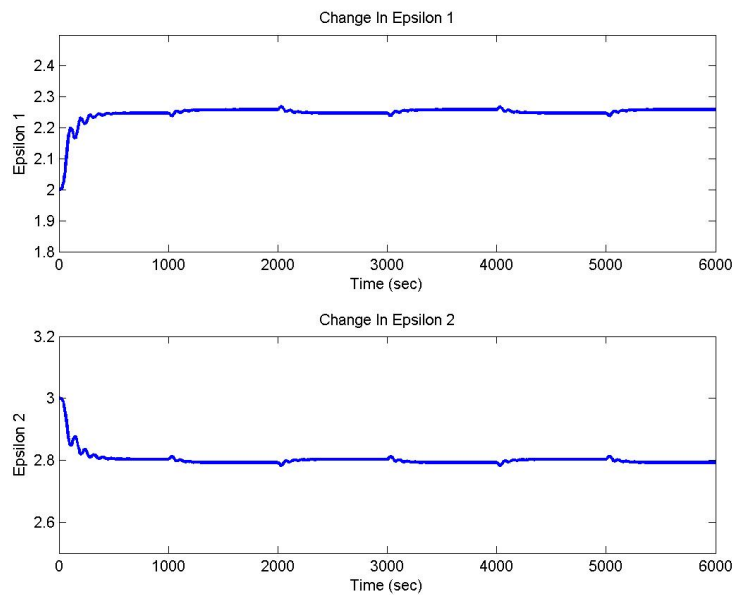


Figure 4.13 – Change in ϵ_1 And ϵ_2 As A Result Of Pulse Command Signal Around $H = 75\%$ With $Q = 50\%$

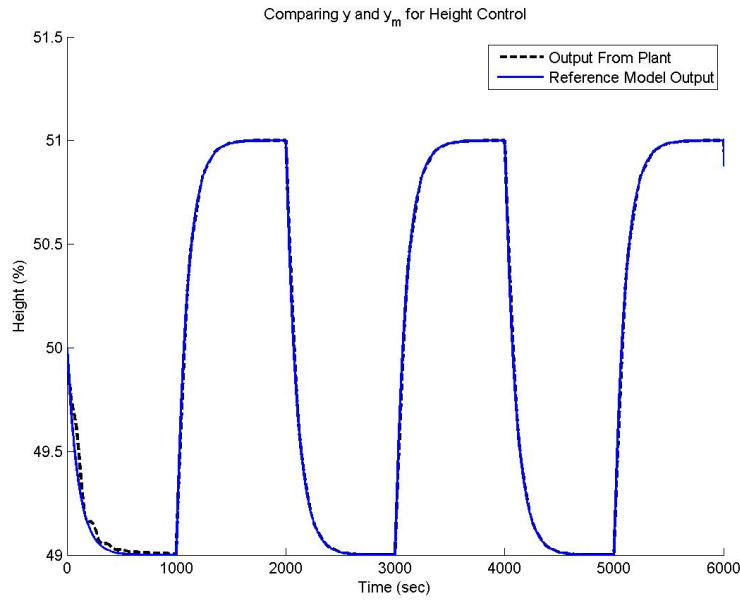


Figure 4.14 – Comparing Outputs From The Plant And The Model Of MRAC With Pulse Command Signal Around $H = 50\%$ With $Q = 60\%$

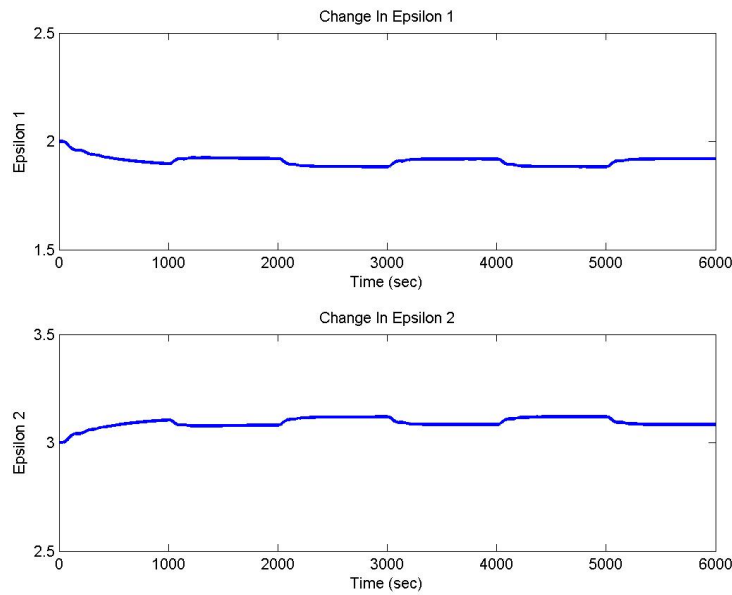


Figure 4.15 – Change in ϵ_1 And ϵ_2 As A Result Of Pulse Command Signal Around $H = 50\%$ With $Q = 60\%$

4.5 Results Of The Height Controllers

The three different controllers (PI, Fuzzy Logic and MRAC) were implemented on the system and the results to set point changes in height are shown in Figure 4.16. The set point of the height of tank two changed from 50% to 55% and then to 60%. The results are shown in Figure 4.16.

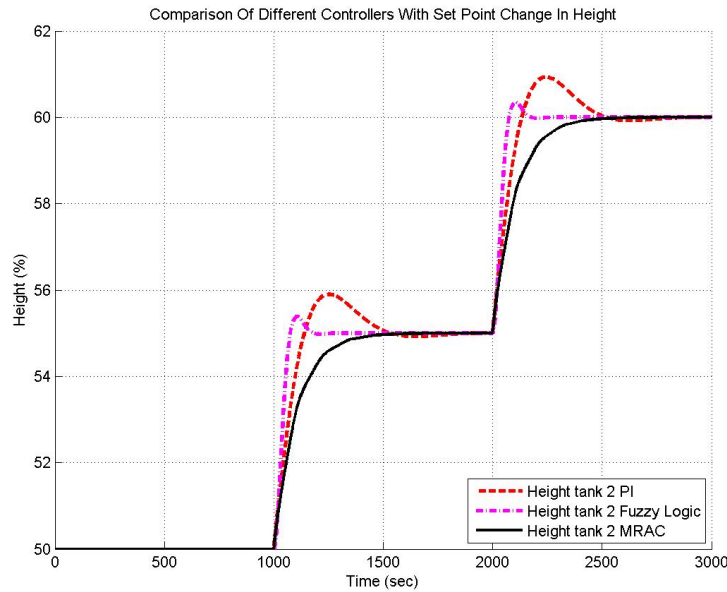


Figure 4.16 – Height Outputs With Change In Set Point, Tank 2

From these results it can be seen that the Fuzzy Logic controller gives the fastest settling time and a very small overshoot of 6%. The PI controller shows an overshoot of 20% and takes longer to settle. The MRAC are over damped and thus have no overshoot.

A change in flow rate has an effect on the heights of the tanks. The next comparison will be to determine which controller will best reject the effect that a change in flow rate has on the height. This is done by changing the input flow rate from 50% to 60%, and back and then to change it from 50% to 40% and back. A sinusoidal disturbance with an amplitude of $0.0005 \text{ m}^3/\text{sec}$ and a frequency of $2\pi \times 0.005 \text{ rad/sec}$ occurs in the input flow rate. The flow rate output from tank one and the height of tank one are shown in 4.17 and 4.18. The outputs from the first tank are the same for all three controllers because the flow from the first tank was designed to follow the filtered value of the input flow. See section 4.1.2. The three different height controllers are implemented on tanks two and three and the results are shown in Figures 4.19 to 4.22. The results of the Fuzzy Logic controller and the MRAC are very similar. All the controllers controlled the heights at 50%, with the PI controller showing the largest deviation from the set point value. The flow rate from the second tank will follow the value of the flow rate from the first tank. The non-linear third sub-plant is prior to the third tank. Therefore

the flow rate from the third tank is lower than the flow rate from the first and second tanks when the input flow rate is lower than 50%.

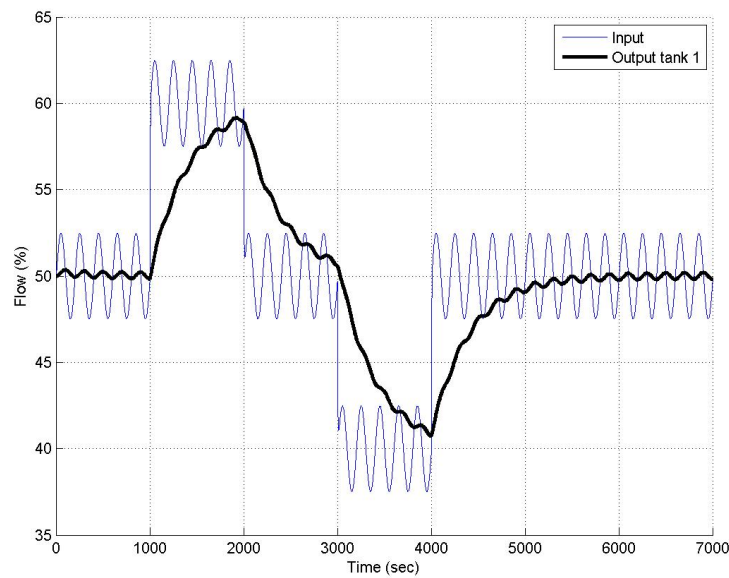


Figure 4.17 – Flow Rate Output From Tank 1

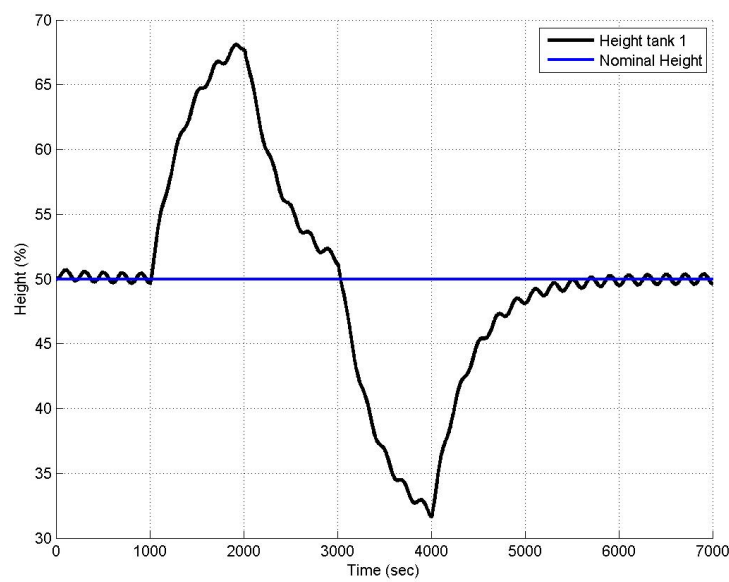


Figure 4.18 – Height Of Tank 1

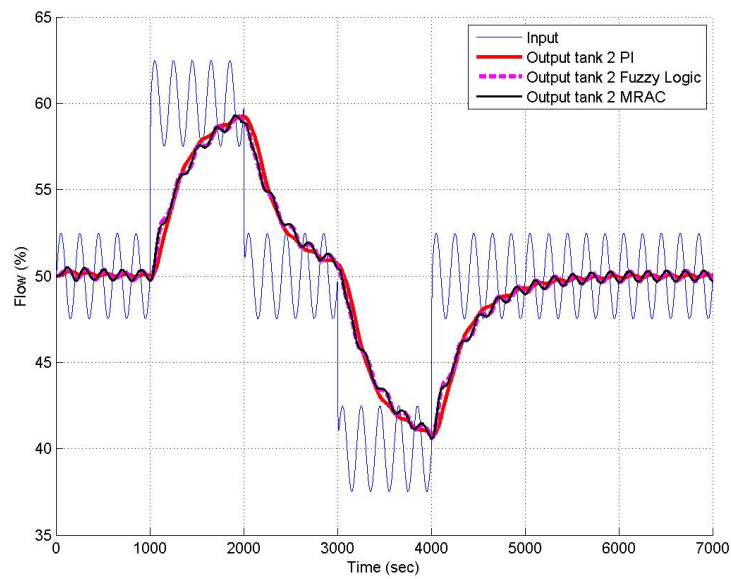


Figure 4.19 – Flow Rate Outputs From Tank 2 With Different Height Controllers

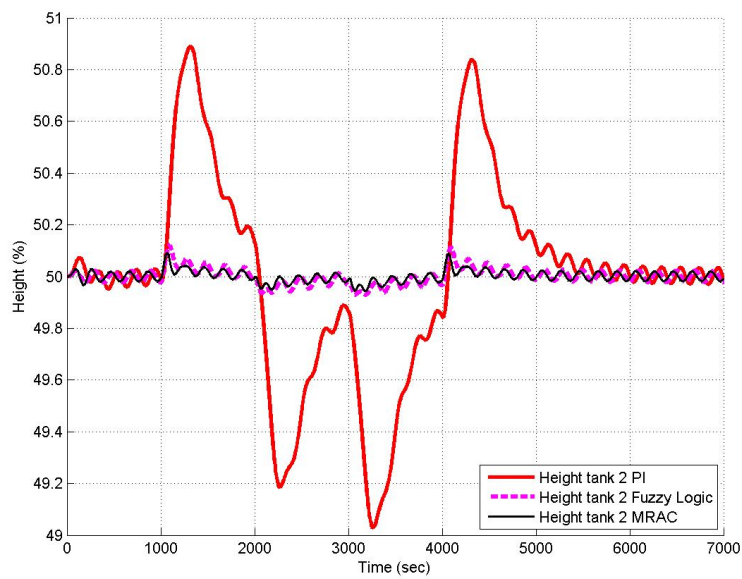


Figure 4.20 – Height Of Tank 2 Different Height Controllers

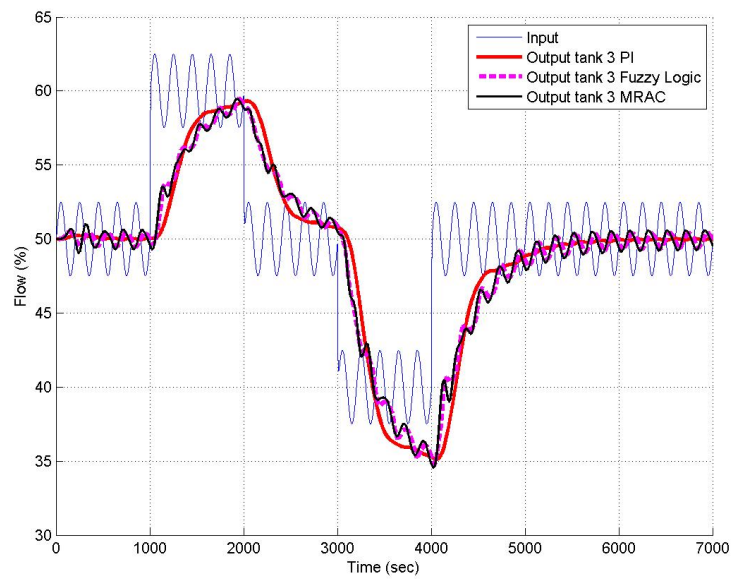


Figure 4.21 – Flow Rate Outputs From Tank 3 With Different Height Controllers

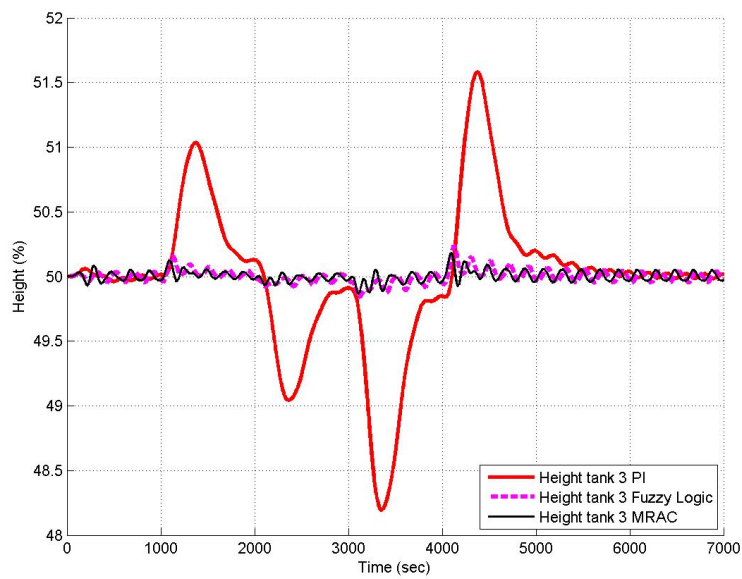


Figure 4.22 – Height Of Tank 3 Different Height Controllers

4.6 Conclusion

The results from the height controllers show that the heights of the liquid levels in the buffers can be controlled, if necessary. The Fuzzy Logic controllers and the MRAC show the best results with the heights of tanks two and three deviating very little from 50%, regardless of disturbances entering the system. The fact that the height is controlled with very little deviation from 50%, the output flow deviates from the nominal throughput because of the input disturbances. Derivative action is added to the Fuzzy Logic controller and if the system is very noisy, the noise can be increased because of this. In this case the PI or MRAC will offer better height control. The height, controlled with the slower PI controller, shows a greater deviation from the set point value, but the flow rate from the tanks shows a smaller effect of the sinusoidal disturbance and is less oscillatory. For efficient disturbance rejection in the flow, though, the buffer capacities should be used and the heights should not be controlled.

The next chapters will discuss the flow control of the system, while keeping the height of the levels in the tanks within boundaries. This will keep the buffers from overflowing or from running empty, but will also make use of the full buffer capacity provided by the tanks.

Chapter 5

Flow Control: Regulatory Controllers

The regulatory flow controllers control the output flow of each sub-plant according to the set point value from the supervisory controller. Four different PI controllers are designed, one for each sub-plant. Then Fuzzy Logic controllers are designed. For the Fuzzy Logic control the same controller design, with different gains, could be used for sub-plants two to four, and another one for sub-plant one. The reason is that sub-plant one does not have a buffer prior to the plant and therefore the controller does not take the height into account. An MRAC is designed for the non-linear sub-plant, sub-plant three. These different control methods are compared.

5.1 Base Case PI Control

To design PI controllers for the four sub-plants, their transfer functions were needed. The transfer functions of the different sub-plants are given in equations 4.2 in section 4.1.1.

To control the output flow from the sub-plants, a valve is used to change the flow rate into the sub-plant. For such a controller to be designed, a transfer function of the change in output flow over the change in valve position is needed. For the first tank, there is no buffer prior to the sub-plant. The output flow from the first sub-plant is controlled with a valve prior to the plant. The flow into the sub-plant is the product of the input stream and the valve position. To generate a transfer function of the change in output flow over the change in valve position, the transfer function of the change in output flow over the change in input flow should be multiplied by the flow rate of the input stream at the work point. The controllers are designed to work nominally at a flow rate of $0.01m^3/sec = 50\%$ and a height of $1m = 50\%$. Thus $G_{p1}(s) = 0.01G_{plant1}(s) = \frac{0.01k_1}{\tau_1 s + 1} e^{-\phi_1 s}$.

There is a buffer for each of the other three sub-plants. The flow rate into the sub-plant is equal to the output flow rate from the tank. The output flow rate from the tank is given by $Q_{out} = 0.01\sqrt{H} \times VP$. The height at the work point at which the function is linearised, is $H = 1m$ and the valve position is $VP = 1$. Therefore the transfer function to control the output flow rate from a sub-plant with the valve from the tank is $G_{pi}(s) = 0.01G_{plant_i}(s) = \frac{0.01k_i}{\tau_i s + 1} e^{-\phi_i s}$, $i=2,3,4$.

These transfer functions can be written in the form:

$$G_{p_i}(s) = \frac{b_i}{s + a_i} e^{-\phi_{p_i}s}, i = 1, 2, 3, 4 \quad (5.1)$$

With

$$\begin{aligned} b_1 &= \frac{0.01}{40} = 0.00025 \\ b_2 &= \frac{0.01}{25} = 0.0004 \\ b_3 &= \frac{0.01k_3}{100} = 0.0001k_3 \\ b_4 &= \frac{0.01}{50} = 0.0002 \end{aligned}$$

$$\begin{aligned} a_1 &= \frac{1}{40} = 0.025 \\ a_2 &= \frac{1}{25} = 0.04 \\ a_3 &= \frac{1}{100} = 0.01 \\ a_4 &= \frac{1}{50} = 0.02 \end{aligned}$$

$$\begin{aligned} \phi_{p1} &= 10 \text{ sec} \\ \phi_{p2} &= 25 \text{ sec} \\ \phi_{p3} &= 5 \text{ sec} \\ \phi_{p4} &= 10 \text{ sec} \end{aligned}$$

The same method as described in section 4.1.2 for the design of the PI controllers is used. The PI controllers are designed to satisfy requirements such as the maximum overshoot, damping ratio, ζ and a specific 2% closed loop settling time, $2\%t_s$. The time delays in the sub-plants add overshoot to the responses. To maintain a small overshoot, the closed loop settling time is chosen long enough to compensate for the effect of the dead-time. For sub-plants one and four, the closed loop 2% settling times are chosen equal to the open loop settling times. Sub-plant three has a relatively small dead-time compared to its time constant and the closed loop 2% settling time is chosen as 300 sec, while the open loop settling time is 400 sec. Sub-plant two has a very large dead time, equal to the plant time constant. This dead time has a big effect on the closed loop response. The damping ratio of sub-plant two is chosen as 1, for closed loop stability. The closed loop settling time is set to 200 sec, while the open loop settling time of the sub-plant is 100 sec. Although this is much slower than the open loop settling time, the response has a small overshoot and the settling time is close to the settling time for sub-plant two with a Fuzzy Logic controller. This is shown in section 5.4. The dominant poles of the closed loop system are a complex conjugate pair with $\zeta < 1$ and repeated and real with $\zeta = 1$. The natural frequency is given by $\omega_n = \frac{4}{\zeta(2\%t_s)}$. To determine ω_n for each sub-plant, the required settling time, $2\%t_s$ is used. Table 5.1 gives the time

delay (ϕ), time constant (τ), open loop 2% settling time (OL 2% t_s), closed loop settling time (CL 2% t_s) as well as the natural frequency (ω_n) and damping ratio (ζ) for each sub-plant.

Plant	ϕ (sec)	τ (sec)	OL 2% t_s (sec)	CL 2% t_s (sec)	ω_n (rad/sec)	ζ
1	10	40	160	160	0.0354	0.707
2	25	25	100	200	0.02	1
3	5	100	400	300	0.0189	0.707
4	10	50	200	200	0.0283	0.707

Table 5.1 – Values Of Different Parameters For Sub-Plants

The transfer function of the PI controller is of the form

$$G_c(s) = \frac{K_P s + K_I}{s} \quad (5.2)$$

with K_P and K_I the proportional and integral gains. The closed loop transfer function is given by

$$T(s) = \frac{G_c(s)G_{pi}(s)}{1 + G_c(s)G_{pi}(s)} \quad (5.3)$$

When the dead time of the plant transfer function is approximated using $e^{-\phi s} = (1 - \phi s)$ [23] and equations 5.1 and 5.2 are substituted in 5.3:

$$\begin{aligned} T_i(s) &= \frac{b_i(1 - \phi_{pi}s)(K_P s + K_I)}{s(s + a_i) + b_i(K_P s + K_I)(1 - \phi_{pi}s)} \\ &= \frac{-K_P b_i \phi_{pi} s^2 + (b_i K_P - K_I b_i \phi_{pi})s + b_i K_I}{(1 - b_i \phi_{pi} K_P)s^2 + (a_i + b_i K_P - K_I b_i \phi_{pi})s + b_i K_I} \end{aligned} \quad (5.4)$$

The denominator of $T_i(s)$ is set equal to $s^2 + 2\zeta\omega_n s + \omega_n^2$, then

$$\frac{a_i + b_i K_P - b_i \phi_{pi} K_I}{1 - b_i \phi_{pi} K_P} = 2\zeta\omega_n \quad (5.5)$$

and

$$\frac{b_i K_I}{1 - b_i \phi_{pi} K_P} = \omega_n^2 \quad (5.6)$$

Equations 5.5 and 5.6 are used to solve for K_P and K_I for each of the four sub-plants. Table 5.2 shows all the parameters for the controllers. The gain of the third sub-plant is at the nominal value with $k_3 = 1$.

Plant	ω_n	ζ	ϕ	a	b	K_P	K_I
1	0.0354	0.707	10	0.025	0.00025	92.47	3.845
2	0.02	1	25	0.04	0.0004	0.2	1
3	0.0189	0.707	5	0.01	0.0001 k_3	158.3	3.275
4	0.0283	0.707	10	0.02	0.0002	94.663	3.243

Table 5.2 – PI Parameters For Sub-Plants

These parameters are used to build PI flow controllers for all four sub-plants. Figure 5.1 shows the implementation of the PI parameters in the PI controller block diagram.

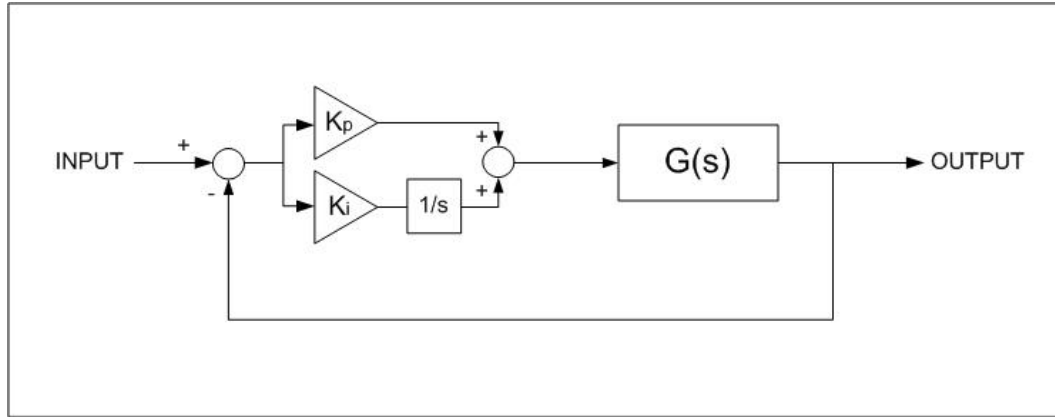


Figure 5.1 – Diagram Of Proportional and Integral Control

These PI controllers are implemented on the cascaded plant system and the results are shown and discussed in section 5.4.

5.2 Fuzzy Logic Control

For the Fuzzy Logic regulatory control, four Fuzzy Logic Controllers are designed to control the flow rates from sub-plants one to four. Mamdani type Fuzzy Logic controllers are used. The different types of Fuzzy Logic controllers are discussed in section 2.2. The input- and output membership functions of a Mamdani Fuzzy Logic controller are given by fuzzy variables. For the Fuzzy Logic control of sub-plants two to four, the same design is used and only the gain values differed. For the control of the flow rate from sub-plant one, a different design is used, because this sub-plant does not have a buffer prior to the plant.

Plants 2-4 To control the flow rates out of the sub-plants, two inputs are used. The first is the flow rate error signal, eQ , given by the difference between the flow rate set point and the output flow rate ($eQ = Q_{SP} - Q_{out}$). The second input is the height of the level of the liquid in the tank (H) prior to the plant. The output from the controller is the change in valve position ($cValve$).

Plant 1 For the Fuzzy Logic controller of sub-plant one, only one input is used. The reason is that there is no buffer to consider prior to the first sub-plant. The only input is eQ , to generate the output, $cValve$. The design of the Fuzzy Logic controller for sub-plant one is similar to the Fuzzy Logic controllers of the other three sub-plants, when the height is at the nominal value of 50%. The design of the Fuzzy Logic controller for sub-plants two to four will first be discussed in section 5.2.1. Subsequently the design of the Fuzzy Logic controller for the first sub-plant will be discussed in section 5.2.2.

5.2.1 Fuzzy Logic Regulatory Control For Sub-Plants 2-4

Input 1, eQ : This input is the difference between the set point for the flow rate Q_{SP} and the measured output flow from the sub-plant, Q_{out} . The set point is an output from

the Supervisory Controller. This difference ($Q_{SP} - Q_{out}$) should be controlled to go to zero, for the output flow rate to follow the given set point value. This is similar to the general Feedback Control principle, shown in Figure 5.2.

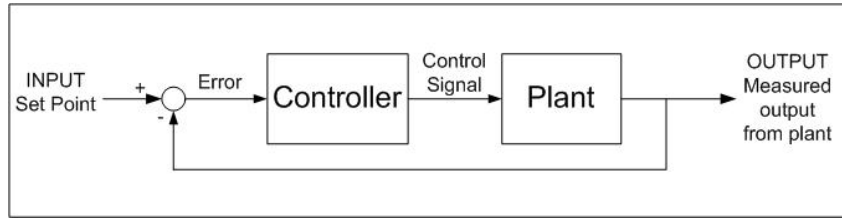


Figure 5.2 – Diagram Of Feedback Control

The membership functions of this input are chosen using the TIL recommendations [5]. These recommendations are described in section 2.2. They recommend to start with three triangular membership functions with an overlap of 50%. One should then decide whether more membership functions are necessary or if some of the functions need to be changed. Three triangular membership functions are sufficient and they are *Negative*, *Zero* and *Positive*. The range of these membership functions are $[-1 \ 1]$ and they are shown in Figure 5.3.

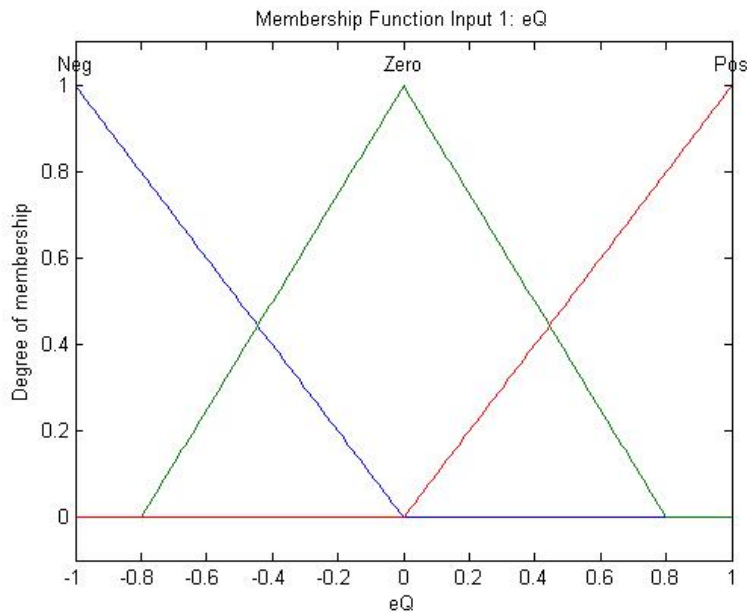


Figure 5.3 – Membership Functions For Input 1: eQ

Negative: $Q_{out} > Q_{SP}$, therefore the output flow needs to decrease. This can be done if the valve opening decreases, thus the change in valve position ($cValve$) is negative.
Zero: $Q_{out} = Q_{SP}$, which means the error signal is zero and therefore $cValve$ is zero.
Positive: $Q_{out} < Q_{SP}$. In this case the output flow rate needs to increase and $cValve$ is positive.

Input 2, H : Because of the height restrictions on the tanks (buffers), the heights should be taken into account to prevent the tanks from overflowing or from running empty. Five membership functions are used to describe H . The reason for this is that if only three membership functions are used, the tank could only either be empty or medium or full. This does not provide enough information. We need to know when the tank is medium empty or medium full. These conditions represent the conditions where the level is still at a safe value, but it is not at the nominal (medium) condition. The membership functions are thus chosen as *Empty*, *Medium Empty*, *Medium*, *Medium Full* and *Full*. These functions are shown in Figure 5.4.

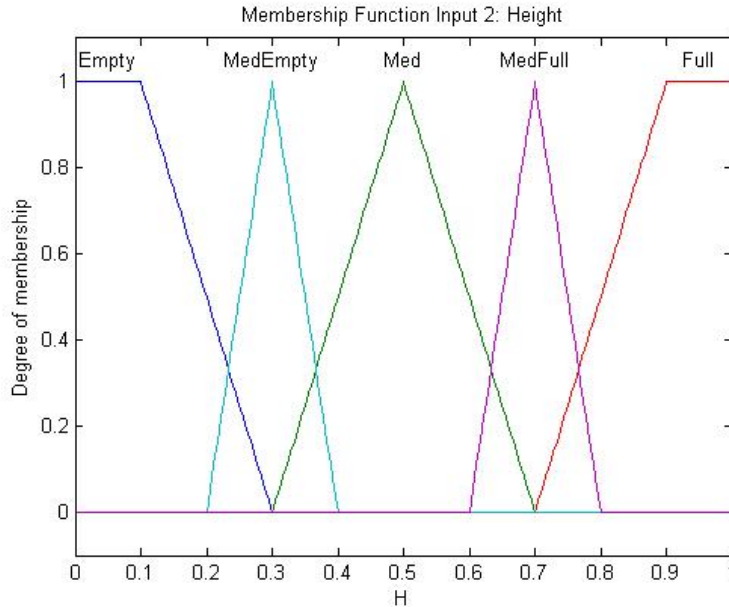


Figure 5.4 – Membership Functions For Input 2: H

The inputs, H and eQ are used to decide how fast and in which direction the valve position should change. For example, the higher the level gets, the slower the valve will close (if at all) and the faster the valve will open. On the other hand, the lower the height, the slower the valve will open (if at all) and the faster it will close. These rules are shown in Table 5.3.

Output 1, $cValve$: The output is used to change the output flow (Q_{out}) by changing the valve position (VP). The range of the membership functions of the output is $[-1 \ 1]$. If $cValve < 0$, VP will become smaller, and therefore decrease Q_{out} . If $cValve > 0$, VP will increase and therefore increase Q_{out} . Five different membership functions are used. They are *Big Negative*, *Small Negative*, *Zero*, *Small Positive* and *Big Positive*. These membership functions not only change the valve position by opening or closing it, but can also change the speed of opening or closing the valve. The membership functions for the output, $cValve$, are shown in Figure 5.5. The rules used to control the outputs of sub-plants two to four, are shown in table 5.3.

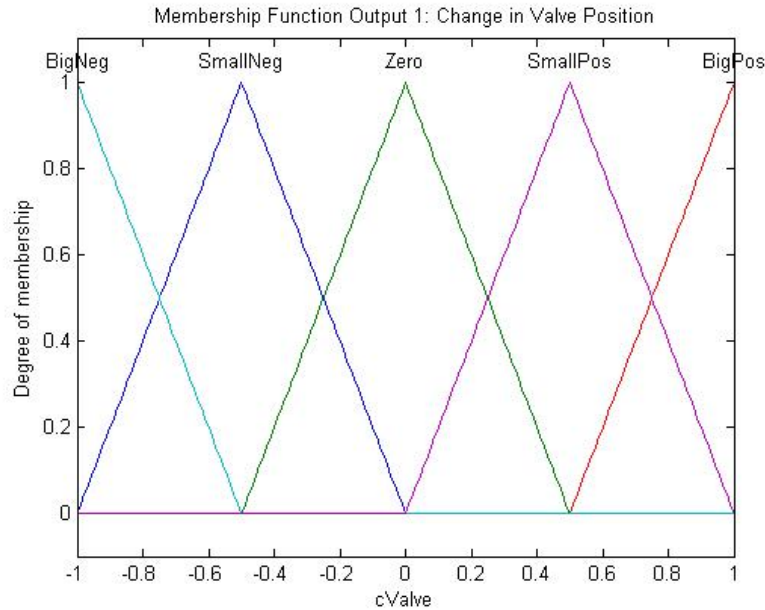


Figure 5.5 – Membership Functions For Output 1: $cValve$

Height	eQ	$cValve$
Empty	Zero	Small Negative
Medium Empty		Zero
Medium		Zero
Medium Full		Zero
Full		Small Positive
Empty	Negative	Big Negative
Medium Empty		Big Negative
Medium		Small Negative
Medium Full		Small Negative
Full		Zero
Empty	Positive	Zero
Medium Empty		Small Positive
Medium		Small Positive
Medium Full		Big Positive
Full		Big Positive

Table 5.3 – Fuzzy Logic Rules For Regulatory Controller

The inputs and output are representing fuzzy variables. The range of these variables are either $[-1 \ 1]$ (eQ and $cValve$) or $[0 \ 1]$ (H). This is not the range of the variables used in each sub-plant. Therefore the values coming from the sub-plant should be normalised before being used in the fuzzy controller. This is called the fuzzification of the input variables. The output should be converted into a value that can be used by the sub-plant. This is called defuzzification.

The range of eQ is $[-1 \ 1]$, where -1 is a big difference between Q_{SP} and Q_{out} , where Q_{out} is bigger than Q_{SP} . A value of 1 is an equally big difference, but with Q_{out} small-

ler than Q_{SP} . The values of Q_{SP} and Q_{out} from the sub-plant are a percentage (%). Therefore the difference has a range of $[-100\% \ 100\%]$. Q_{SP} is not expected to differ from Q_{out} by 100%, but a value of 10% is a big difference. The value of a big difference in the fuzzy controller $[-1 \ 1]$ should represent $[-10\% \ 10\%]$. The fuzzification gains of eQ is $FG_Q = 0.1$ for sub-plants two to four. The height value from the sub-plant is a percentage (%) and should be in the range $[0 \ 1]$. The whole range (0% – 100%) is used. Therefore the fuzzification gain for H is $FG_H = 0.01$.

The defuzzification gain of the change in valve position, DG_{VP} , determines the speed at which the valve position changes. If it is too big, the change will be too aggressive and the output can be oscillatory. On the other hand, if it is too small, the valve position will change very slowly. The transfer function that describes the change in VP with a change in Q_{SP} , is non-linear and tested around the nominal value of 50%, as well as at values away from the nominal flow rate. The defuzzification gains were chosen $DG_{VP2} = 0.004$ for sub-plant two, $DG_{VP3} = 0.003$ for sub-plant three and $DG_{VP4} = 0.0035$ for sub-plant four. These values depended on the different dynamics of the sub-plants and are chosen to create the fastest response with a small overshoot ($\approx 10\%$). The controller performances can be seen in section 5.4. Figure 5.6 shows a diagram of the Fuzzy Logic regulatory controller.

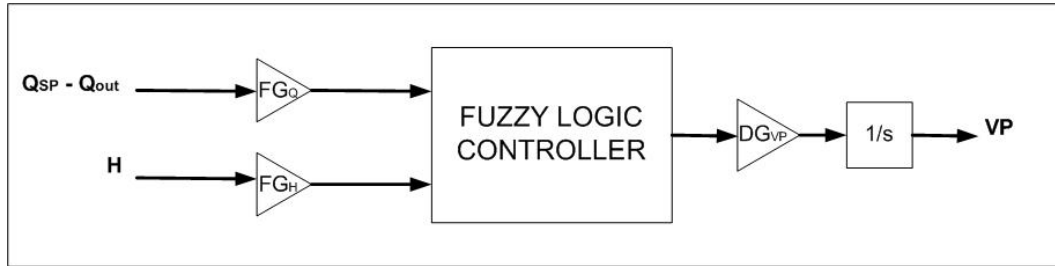


Figure 5.6 – Diagram Of Fuzzy Logic Controller For Regulatory Control

5.2.2 Fuzzy Logic Regulatory Control For Sub-Plant 1

Input 1, eQ : This input is the only input and represents the difference between the set point value for the flow rate Q_{SP} and the measured output flow from the sub-plant Q_{out} . As for the Fuzzy Logic controller for sub-plants two to four, this input could be represented by membership functions *Negative*, *Zero* and *Positive*, but the fact that this is the only input, means this is the only information used to generate an output, $cValve$. To generate $cValve$ with more than three membership functions, two extra membership functions were added to eQ . They are *Small Negative* and *Small Positive*. This makes more sensitive control of the output flow rate possible. These membership functions for $cValve$ are the same as for the controller for sub-plants two to four. The membership functions for eQ can be seen in Figure 5.7. The fuzzification gain for eQ is 0.1 and the defuzzification gain for $cValve$ is 0.0035. A defuzzification gain of 0.0035 gives an overshoot of less than 10%.

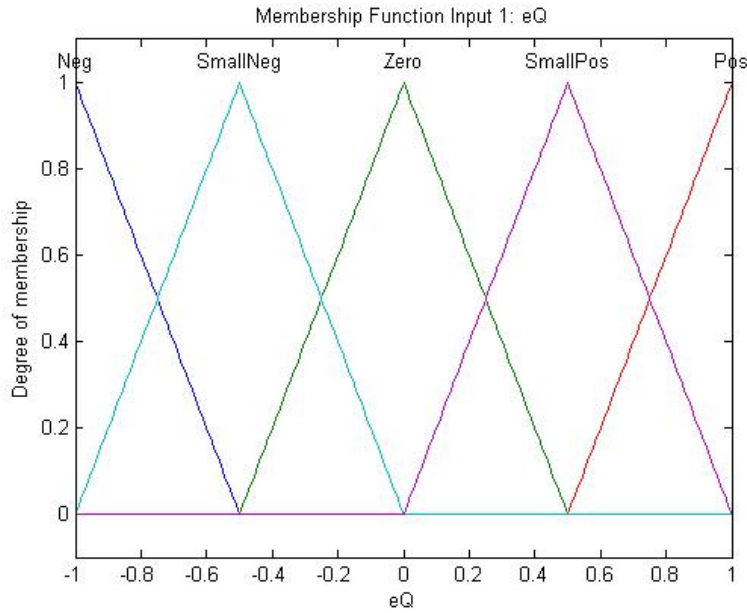


Figure 5.7 – Membership Functions For Input 1: eQ

The rules for the control of the output flow of sub-plant one, are shown in table 5.4. The Fuzzy Logic controllers are compared with the other controllers in section 5.4.

eQ	$cValve$
Negative	Big Negative
Small Negative	Small Negative
Zero	Zero
Small Positive	Small Positive
Positive	Big Positive

Table 5.4 – Fuzzy Logic Rules For Regulatory Controller For Sub-Plant 1

5.3 Model Reference Adaptive Control For The Non-linear Sub-Plant 3

The third sub-plant has a non-linear transfer function. The gain differs for different input flow rates. Therefore a Model Reference Adaptive Controller is designed to control the output from the third sub-plant. An adaptive controller can adapt the controller parameters with varying plant gain.

When designing an MRAC, the following needs to be selected: A reference model, the controlling structure and tuning gains. The controller structure seen in Figure 5.8 are used. It is based on the MIT rule, discussed in section 2.3. The model chosen to be followed is of the form $\frac{a_m}{s+a_m}e^{-\phi_m s}$. This model was chosen as a first order function with a settling time of 200 sec and a dead time similar to open loop model of $\phi_m = \phi_{p3} = 5$ sec.

The time constant is given by $4 \times \tau_m = 2\%t_s = 200 \text{ sec}$. Therefore $\tau_m = 50 \text{ sec} = \frac{1}{a_m}$. Thus $a_m = 0.02$ and $G_m(s) = \frac{0.02}{s+0.02} e^{-5s}$.

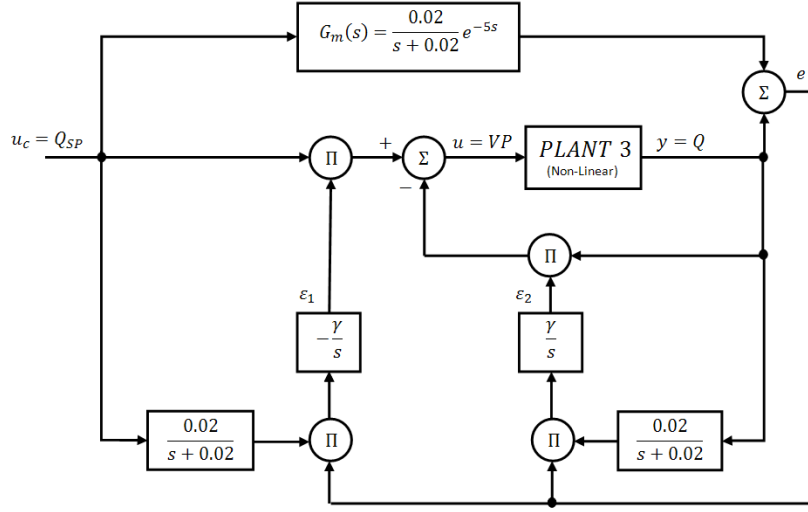


Figure 5.8 – Block Diagram Of The Model Reference Adaptive Controller For Sub-Plant 3

The initial values of ϵ_1 and ϵ_2 were determined from the diagram in Figure 5.8, which is simplified in Figure 5.9. The transfer function of sub-plant 3, from equation 5.1, is $G_{p3}(s) = \frac{b_3}{s+a_3} e^{-\phi_{p3}s} = \frac{0.0001k_3}{s+0.01} e^{-5s}$. The delay is approximated by $e^{-5s} \approx (1-5s)$, [23], p.116. This yields

$$\begin{aligned} \frac{y}{u_c} &= \frac{G_{p3}(s)\epsilon_1}{1 + G_{p3}(s)\epsilon_2} \\ &= \frac{0.0001k_3\epsilon_1 e^{-5s}}{s + 0.01 + 0.0001k_3\epsilon_2 e^{-5s}} \\ &\approx \frac{\frac{0.0001k_3\epsilon_1}{1-0.0005k_3\epsilon_2}}{s + \frac{0.01+0.0001k_3\epsilon_2}{1-0.0005k_3\epsilon_2}} e^{-5s} \end{aligned} \quad (5.7)$$

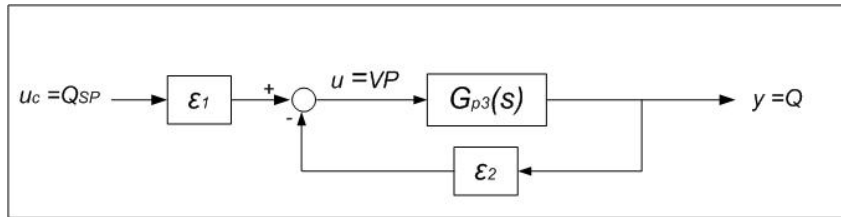


Figure 5.9 – Model Reference Adaptive Control Diagram To Calculate The Initial Values For ϵ_1 and ϵ_2

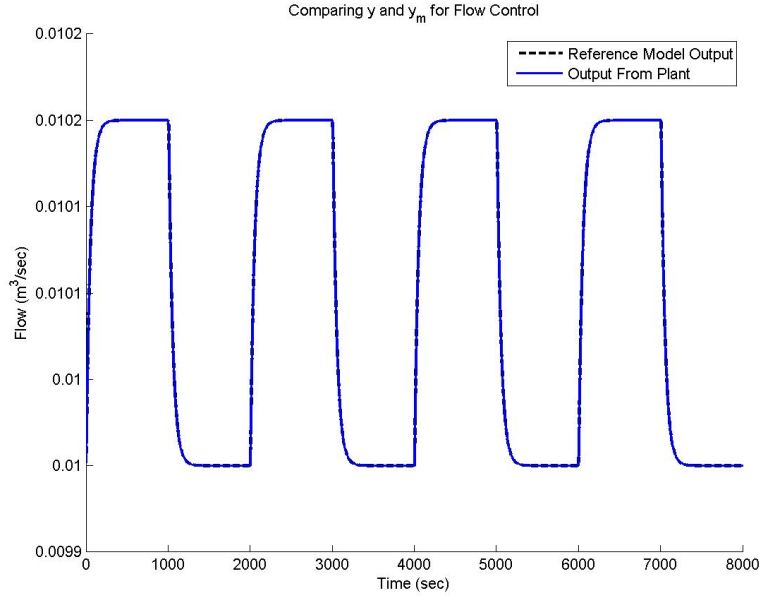


Figure 5.10 – Outputs Of Plant And Model For MRAC With $\gamma = 10000$, ($Q = 50\% = 0.01 \text{ m}^3/\text{sec}$ and $k_3 = 1$)

To determine initial values for ϵ_1 and ϵ_2 , the equation $\frac{y}{u_c} = \frac{y_m}{u_c}$ is solved. This means

$$\frac{\frac{0.0001k_3\epsilon_1}{1-0.0005k_3\epsilon_2}}{s + \frac{0.01+0.0001k_3\epsilon_2}{1-0.0005k_3\epsilon_2}} e^{-5s} = \frac{0.02}{s + 0.02} e^{-5s} \quad (5.8)$$

which yields $\epsilon_1 = 190.91$ and $\epsilon_2 = 90.91$ at the nominal work points of $Q = 0.01 \text{ m}^3/\text{sec}$ where $k_3 = 1$. These values also satisfy equation 5.9 from Figure 5.9, with nominal conditions of $u_c = Q_{SP} = 0.01 \text{ m}^3/\text{sec}$, $y = Q_{out} = 0.01 \text{ m}^3/\text{sec}$ and $u = VP = 1$:

$$\begin{aligned} \epsilon_1 u_c - \epsilon_2 y &= u \\ 0.01\epsilon_1 - 0.01\epsilon_2 &= 1 \\ \epsilon_1 - \epsilon_2 &= 100 \end{aligned} \quad (5.9)$$

The next step in the design of the MRAC is to determine the value of γ . A square wave disturbance is added to the set point flow rate and the outputs are measured. When the sub-plant operates in the non-linear area, the controller parameters change to new values. The rate at which these parameters adapt is determined by the value of γ . The results from the step-inputs are shown in Figures 5.10 to 5.13 at different flow rates. To change the values of the controller parameters within 1000 sec, the value of γ was set to 10000. This is shown in Figure 5.14.

The MRAC can control the output flow rate of the non-linear sub-plant at different flow rates. When the flow rate is equal to $0.01 \text{ m}^3/\text{sec}$, the output from the plant follows the output from the model without any error as shown in Figure 5.10. At flow rates greater than $0.01 \text{ m}^3/\text{sec}$, the controller parameters change to new values and then

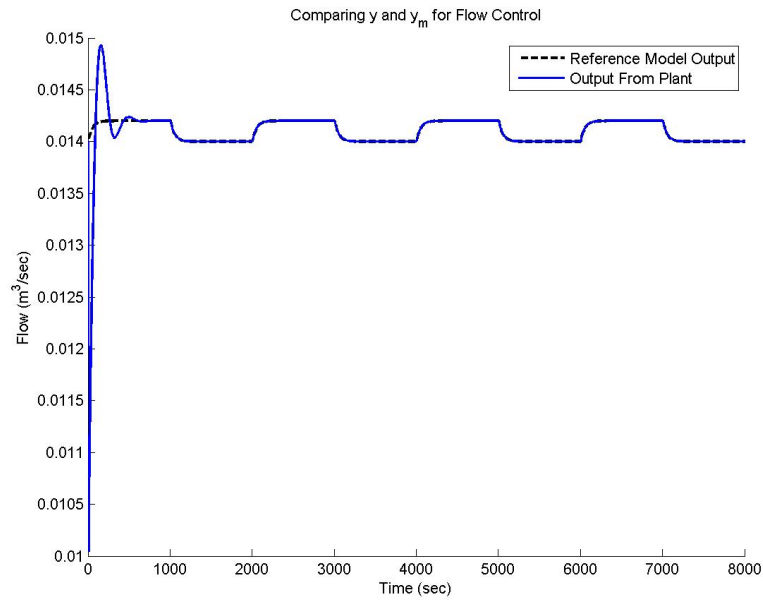


Figure 5.11 – Outputs Of Plant And Model For MRAC With $\gamma = 10000$, ($Q = 70\% = 0.014 \text{ m}^3/\text{sec}$)

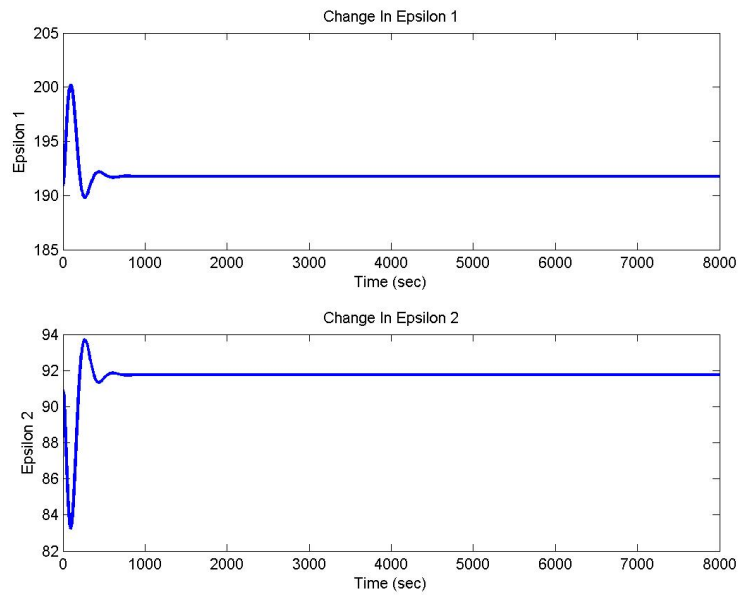


Figure 5.12 – Change In ϵ_1 And ϵ_2 With $\gamma = 10000$, ($Q = 70\% = 0.014 \text{ m}^3/\text{sec}$)

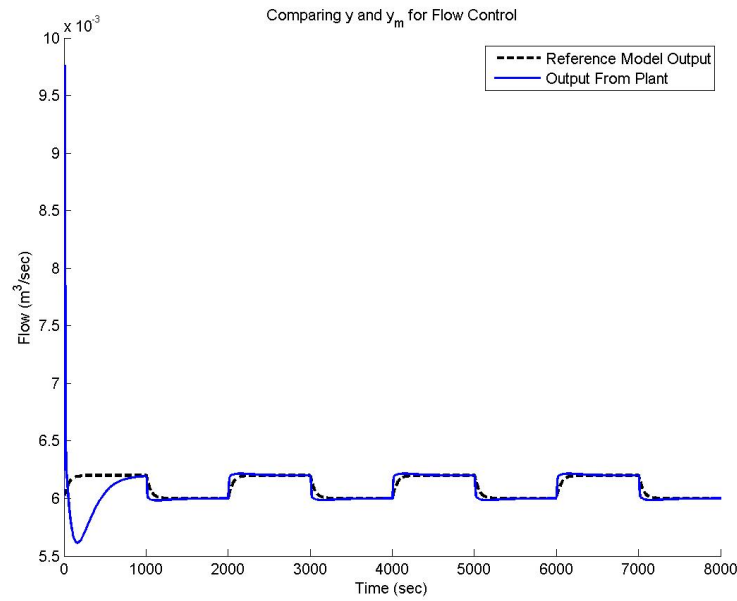


Figure 5.13 – Outputs Of Plant And Model For MRAC With $\gamma = 10000$, ($Q = 30\% = 0.006 \text{ m}^3/\text{sec}$)

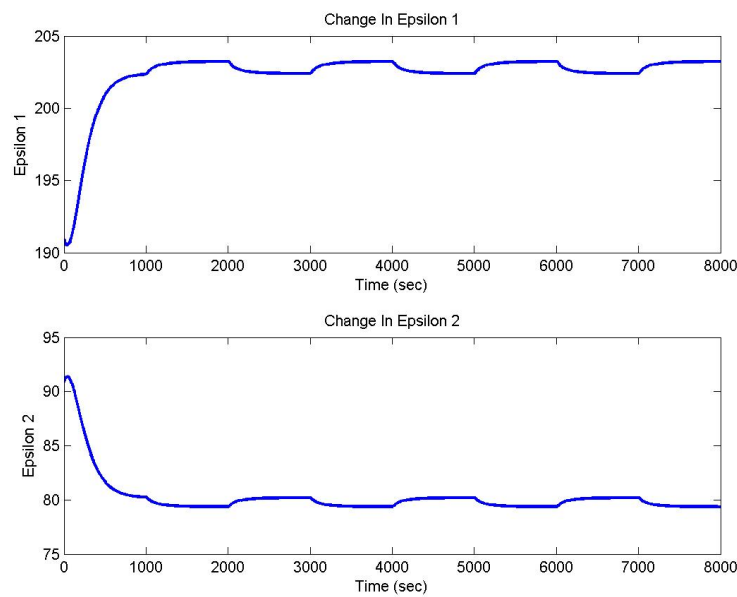


Figure 5.14 – Change In ϵ_1 And ϵ_2 With $\gamma = 10000$, ($Q = 30\% = 0.006 \text{ m}^3/\text{sec}$)

stays constant, Figure 5.12. When the new parameter values are reached, the output from the plant follows the model perfectly, Figure 5.11. With a flow rate slower than the nominal flow rate of $0.01 \text{ m}^3/\text{sec}$, the plant gain changes when the flow rate changes. With a flow rate of 30%, the controller parameters adapt to their values for the new flow rate, but each set point change changes the parameters, because the gain of the sub-plant changes each time, Figure 5.14. Therefore there is a slight difference between the output from the plant and the output from the model. This can be seen in Figure 5.13. The error between the outputs of the plant and the model is small with $\gamma = 10000$ and the controller performs satisfactory at flow rates slower than the nominal flow rate.

5.4 Experimental Results

5.4.1 Closed Loop Flow Control

The different controller responses to step set point changes are compared in this section. To compare the responses at different flow rates, a step from the nominal flow rate of $Q = 50\%$ to 60% and then to 70% is implemented. Then a down step from $Q = 50\%$ to 40% and then to 30% is applied. This shows the controller performances at flow rates away from the nominal flow rates. For sub-plants one, two and four, the PI controllers and Fuzzy Logic controllers are compared. In these cases, the response of each controller is the same at different flow rates. In each case either the step up or step down set point changes are shown. For sub-plant three, the PI controller, MRAC and Fuzzy Logic controller are compared. The responses of the non-linear sub-plant are different at different flow rates and both the step up and step down set point changes are shown. The results are shown in Figures 5.15 to 5.19.

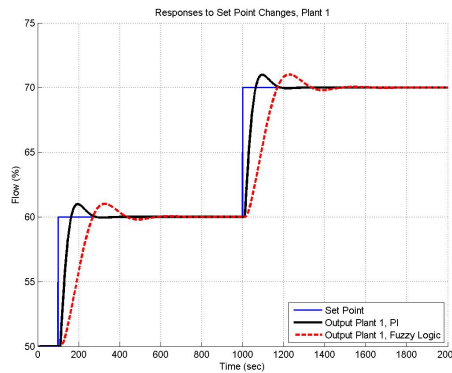


Figure 5.15 – Step Up Responses Of PI And Fuzzy Logic Flow Controller, Sub-Plant 1

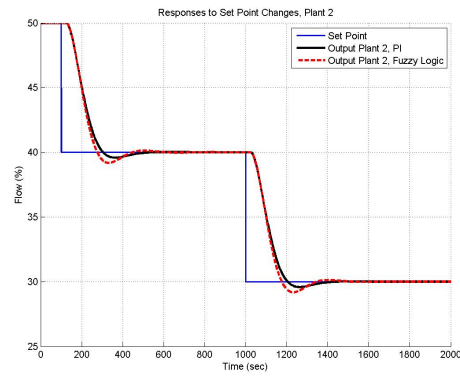


Figure 5.16 – Step Down Responses Of PI And Fuzzy Logic Flow Controller, Sub-Plant 2

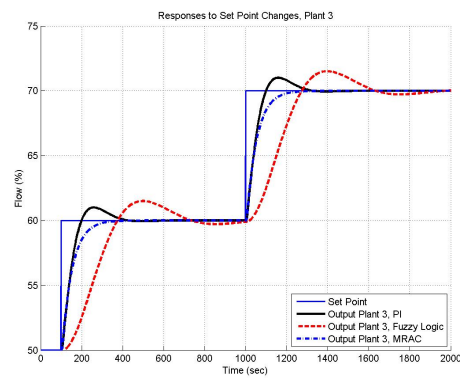


Figure 5.17 – Step Up Responses Of PI And Fuzzy Logic Flow Controller, Sub-Plant 3

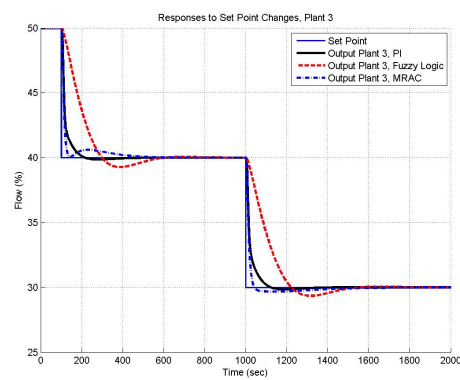


Figure 5.18 – Step Down Responses Of PI And Fuzzy Logic Flow Controller, Sub-Plant 3

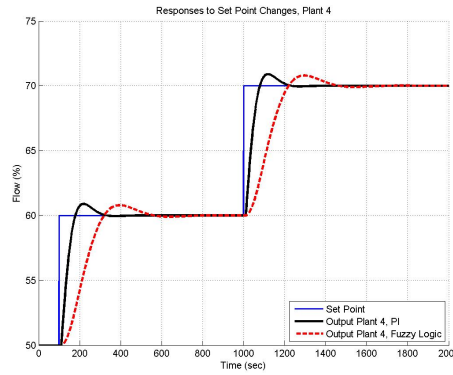


Figure 5.19 – Step Up Responses Of PI And Fuzzy Logic Flow Controller, Sub-Plant 4

For sub-plants one, three and four the Fuzzy Logic controllers show a slower response than the PI controllers or the MRAC on sub-plant three. For sub-plant two the responses from the PI and Fuzzy Logic controllers are very similar. Better disturbance rejection is expected from the faster PI controllers and the MRAC. The disturbance rejection from the Fuzzy Logic controllers as well as from the other controllers, are inspected and the results are shown in section 5.4.2.

5.4.2 Disturbance Rejection During Flow Control

To compare the different regulatory controllers, a sinusoidal disturbance with an amplitude of $0.0005 \text{ m}^3/\text{sec}$ (5% of nominal input of $0.01 \text{ m}^3/\text{sec}$) and a frequency of $2\pi \times 0.005 \text{ rad/sec}$ is added to the input. Then additional step disturbances of $0.002 \text{ m}^3/\text{sec}$ (20% of nominal input) are added as follows: After 1000 sec a step-up and a step-down after 2000 sec. Then another step-down after 3000 sec and a step-up after 4000 sec.

These results are shown in Figures 5.20 to 5.23. The outputs of all the different sub-plants are given in Figure 5.20 for PI control and in Figure 5.22 for Fuzzy Logic control. The outputs from sub-plants three and four are shown individually in Figures 5.21 and 5.23. Here the results with and without the MRAC on sub-plant three can be compared.

From these graphs, it can be seen that both the PI controllers and the Fuzzy Logic controllers reject the disturbances, but the outputs from the PI controllers deviate less from the set point value of 50%. This can be explained by the fact that the PI controllers are faster with shorter settling times. It can be seen that the Fuzzy Logic controller changes less frequently and therefore deviates slightly more from the set point values. The outputs from the last sub-plant are very close to the set point values with both kinds of controllers and therefore the PI control as well as the Fuzzy Logic control is sufficient for disturbance rejection control. When MRAC is added to sub-plant three, it improves disturbance rejection slightly with Fuzzy Logic control, but with PI control the disturbance rejection is worse when a MRAC is added. The statistical evaluation of these results are shown in section 5.4.3.

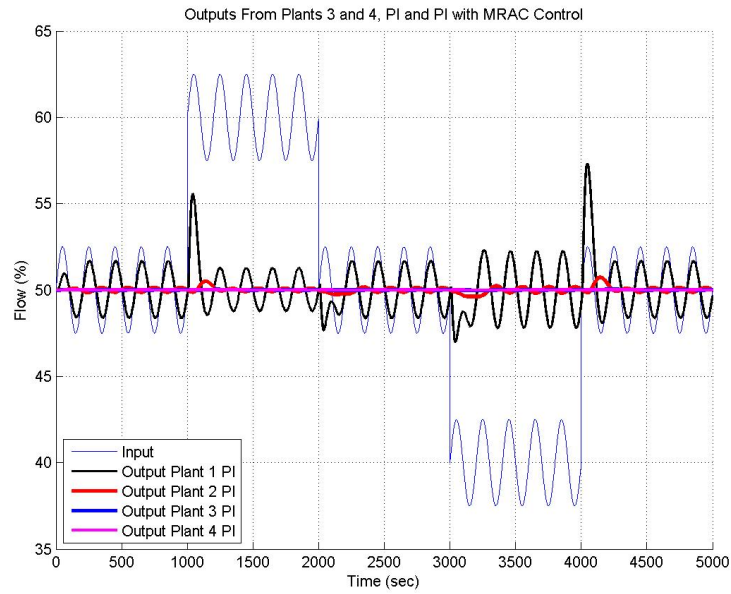


Figure 5.20 – Outputs From Different Sub-Plants With PI Control

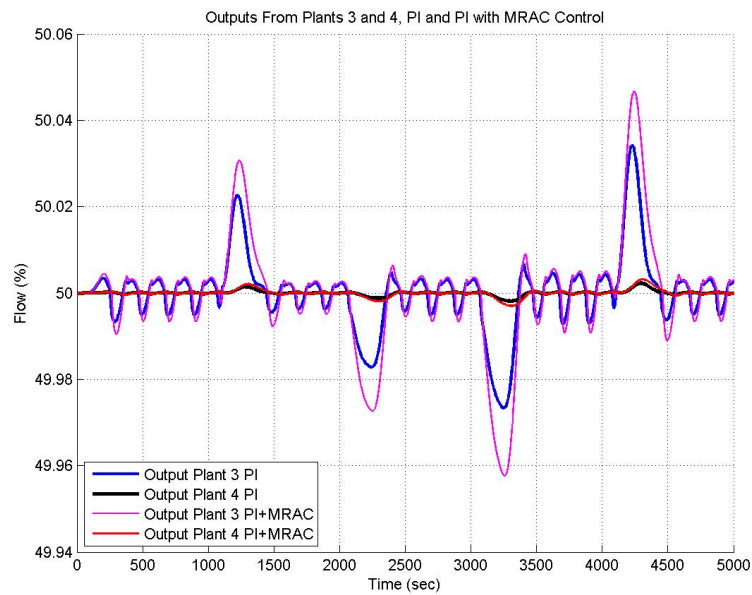


Figure 5.21 – Comparing Outputs From Sub-Plants 3 And 4 With PI Only And PI Plus MRAC Control

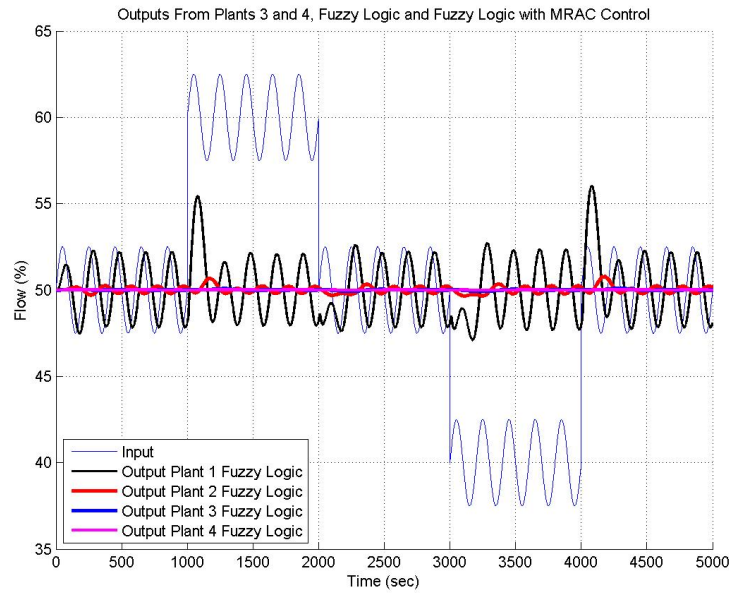


Figure 5.22 – Outputs From Different Sub-Plants With Fuzzy Logic Control

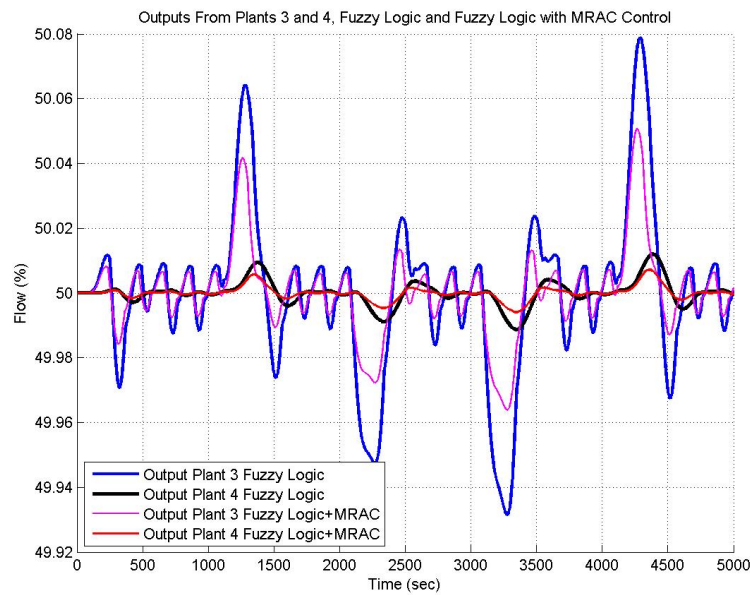


Figure 5.23 – Comparing Outputs From Sub-Plants 3 And 4 With Fuzzy Logic Control And Fuzzy Logic With MRAC On Sub-Plant 3

5.4.3 Statistical Evaluation Of Disturbance Rejection

To compare the controllers, a value is generated that shows how much the input disturbances are rejected. An error signal is determined by the output of each sub-plant minus the nominal value of 50%. The RMS (Root mean square) of the error at every time step is calculated and compared. The RMS value is defined in Equation 5.10 with e_i the error at time step i and n the number of measurements (time steps). The inputs and disturbances are the same as for the graphical results in section 5.4.2.

$$RMS = \sqrt{\frac{e_1^2 + e_2^2 + e_3^2 + e_4^2 + \dots}{n}} \quad (5.10)$$

A frequency analysis of the closed loop response is performed by calculating the RMS values with the same disturbances, but the frequency of the sinusoidal disturbance is changed. The frequency of the sinusoidal disturbance was set to $2\pi \times 0.005$ rad/sec for the graphical comparison between the different controllers. A sinusoidal disturbance with a higher frequency of $2\pi \times 0.008$ rad/sec as well as a sinusoidal disturbance with a lower frequency of $2\pi \times 0.002$ rad/sec is implemented on the system. The results are shown in Table 5.5.

Control Method	Plant 1	Plant 2	Plant 3	Plant 4
$2\pi \times 0.008$ rad/sec				
PI	1.3847	0.1360	0.0081	$5.6e^{-4}$
PI+MRAC	1.3847	0.1360	0.0124	$8.66e^{-4}$
FUZZY	1.299	0.143	0.0227	0.0037
FUZZY+MRAC	1.299	0.143	0.0123	0.002
$2\pi \times 0.005$ rad/sec				
PI	1.506	0.164	0.0085	$5.8e^{-4}$
PI+MRAC	1.506	0.164	0.0128	$8.8e^{-4}$
FUZZY	1.7994	0.205	0.024	0.0039
FUZZY+MRAC	1.7994	0.205	0.0134	0.002
$2\pi \times 0.002$ rad/sec				
PI	1.089	0.1784	0.0109	$7.57e^{-4}$
PI+MRAC	1.089	0.1784	0.0165	0.0011
FUZZY	1.866	0.3044	0.0512	0.0084
FUZZY+MRAC	1.866	0.3044	0.0268	0.0044

Table 5.5 – Root-Mean-Square Errors

From these results it is shown that the PI controllers, with a lower RMS value, deviates less from the set point value than the Fuzzy Logic controllers, but both kinds of controllers have sufficient disturbance rejection. With the MRAC implemented on the third sub-plant, the results with the Fuzzy Logic controllers improve slightly, but the PI controllers reject disturbances better without the MRAC on the third sub-plant.

The results at different frequencies show that the disturbance rejection is sufficient for this range of disturbance frequencies. The added step inputs show the rejection of a slow frequency disturbance.

5.5 Conclusion

The results of the different regulatory controllers are shown in sections 5.4.2 and 5.4.3. From these results the following can be seen. With flow rate set point changes, the PI controllers show a faster response than the Fuzzy Logic controllers in all the sub-plants except in sub-plant two. The settling times of the PI and Fuzzy Logic controllers for sub-plant two are very similar. The control of the non-linear sub-plant (sub-plant three) is done by a PI controller, a Fuzzy Logic controller and a MRAC. The MRAC and PI controllers show similar settling times with a step up in set point value. The PI controller has a small overshoot while the MRAC is over damped with no overshoot. In this case, the sub-plant has a constant gain of 1. With a step down in set point value, the gain of the sub-plant changes and the MRAC shows a faster response than the PI controller. In both the step up and step down set point changes, the Fuzzy Logic controller has the slowest response to ensure closed loop stability.

The PI and Fuzzy Logic regulatory controllers reject disturbances in the input and the addition of a MRAC to sub-plant three does not change the results of these controllers very much. With PI control the addition of the MRAC makes the disturbance rejection vaguely worse, but with Fuzzy Logic control the MRAC on sub-plant three improves the disturbance rejection slightly. The different types of disturbances used are sinusoidal and pulse disturbances of different frequencies. The PI controllers show a smaller deviation from the set point values than the Fuzzy Logic controllers, but even with a slower response, the Fuzzy Logic controllers still show very good disturbance rejection. The different regulatory controllers are implemented under a supervisory controller, described in the next chapter (Chapter 6).

Chapter 6

Flow Control: Supervisory Controller

Supervisory Control is finally implemented on the system. The supervisory controller is the outer layer of the two layers of control, refer to section 2.5. The inner layer consists of regulatory controllers that control the output flow from each tank. The regulatory controllers do not control the level of the liquid in each tank to a certain reference height. To prevent the system from overflowing or running dry, safety controllers will be added to the inner layer. The Fuzzy Logic regulatory controllers do take the height into account when determining the output valve positions. The Fuzzy Logic regulatory controllers can keep the height within safe boundaries while controlling the output flow when changing the valve position. Therefore the safety controllers must be used in abnormal situations. The set points for the output flow rates are calculated in the supervisory controller. These set points will not only depend on the input flow, but also on any capacity restrictions in the sub-plants. The height of the liquid level gives an indication of the current buffer capacity and will be used to determine these set points as well. The safety controllers will be discussed in section 6.1 and the supervisory control will be discussed in section 6.2.

Although it is preferable to keep the output flow rate steady at its nominal value of 50%, it will not be possible due to disturbances and other restrictions in the plant system. This is why the set points for the regulatory flow-controllers cannot be constant at the nominal value, and should be determined elsewhere. The different restrictions are height restrictions on tanks one, two and three, input capacity restrictions and capacity restrictions at sub-plants one, two and four. Disturbances could also occur in the input. These disturbances can be in the form of a sinusoidal disturbance, a step disturbance, a pulse disturbance or any combination of these. The capacity restrictions on the input flow as well as on the different sub-plants can be anything between 0% and 100%, which means a flow rate of anything between $0\text{ m}^3/\text{sec}$ to $0.02\text{ m}^3/\text{sec}$. For example, with any sub-plant having a capacity restriction of 40%, a flow rate of $0.4 \times 0.02\text{ m}^3/\text{sec}$ must be met. If a higher flow enters the sub-plant, the extra amount of liquid will be dumped and lost. This decreases the profit, therefore under such conditions a cutback in throughput should be activated to prevent such losses.

To implement Supervisory Control, the buffer capacity is exploited. With an area, $A = 2\text{ m}^2$ and a maximum height, $H = 2\text{ m}$, each tank has a volume of 4 m^3 . With a nominal flow of $0.01\text{ m}^3/\text{sec}$, it can take a maximum of 400 sec for the water to pass through the tank as buffer.

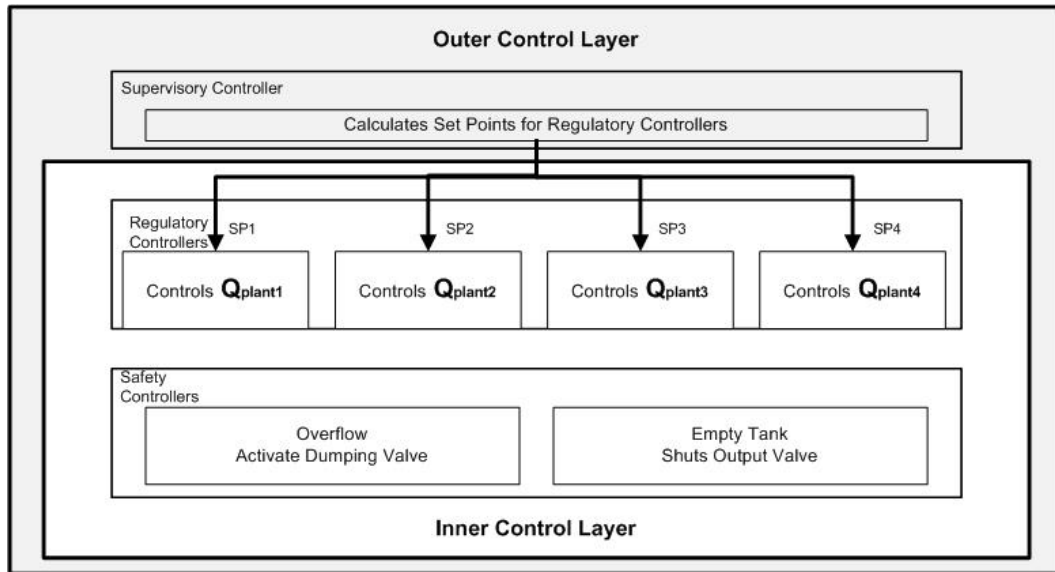


Figure 6.1 – Diagram Of The Different Control Layers Of The System

6.1 Safety Control

Safety controllers are added to the system. These safety controllers operate outside the regulatory or supervisory controllers. It keeps the tanks from overflowing and from running empty. Although this is prevented in the Fuzzy Logic regulatory controllers, an external safety mechanism is implemented in addition. This will be used in abnormal situation management and works as follows: A dump valve exists on all the tanks. This valve is normally shut (dump valve position, $VP_{dmp} = 0\%$). When the level reaches a height of 80%, the valve is opened half way ($VP_{dmp} = 50\%$). This would force the level to a lower height. This valve is closed again ($VP_{dmp} = 0\%$) when the height is below 75%. In an abnormal condition when the height increases very fast and exceeds 90%, the dump valve is opened to its full capacity ($VP_{dmp} = 100\%$). This will prevent the tank from overflowing. The liquid lost through dumping is a loss and will decrease the profit of the whole plant system. Therefore it is not preferable, but it is a safety mechanism that will only be used under abnormal conditions.

The second part of the safety control is the prevention of a tank running empty. This is done by shutting the outflow valve when the level reaches a height lower than 5%. When the outflow valve is closed, and there is still flow into the tank, the height will increase. The output valve will only open again when the level reaches a height greater than 10%. This mechanism is used in the PI regulatory controllers to prevent the tanks from running empty. For Fuzzy Logic regulatory controllers it is used for abnormal situation management because the heights are already taken into account when calculating the valve position in the Fuzzy Logic regulatory controllers, which will normally prevent the tanks from running empty before they reach the extreme value of 5%.

6.2 Supervisory Control

The supervisory controller consists of two parts. The first is used to calculate the lowest capacity in the whole plant system. This value is used in the second part where the set points for the various sub-plant outputs are determined.

6.2.1 Supervisory Control Part 1: To Determine The Cut-back Set Point (SP_{cb})

To determine the lowest throughput capacity in the whole system, the different sub-plant capacity restrictions and the capacity of the input source are compared. The capacity restrictions used are the maximum input flow rate as well as the capacity restrictions of sub-plants one, three and four. No capacity restrictions occur in sub-plant two, because no external substances are used for the process of sub-plant two. The output of this comparison between throughput capacities is the lowest overall capacity at each time step. This is called the cutback set point (SP_{cb}) and is an input to the next part of the supervisory controller. The values of the capacity restrictions is used, but no indication of where the restrictions occur is necessary to determine SP_{cb} . The location of each restrictions is used in the second part of the supervisory controller and is entered as an input, *Throughput Capacity, TC*.

6.2.2 Supervisory Control Part 2: To Determine The Set Points For The Regulatory Controllers

For the supervisory control, a Sugeno type Fuzzy Logic controller is used. With a Sugeno Fuzzy Logic controller, the output membership functions are not a set of fuzzy values, but the output membership functions are functions of the inputs.

The inputs to the supervisory controller are:

- The maximum input flow rate from the source into the system, *InputFlow*
- The cut-back set point from the first part of the supervisory controller, SP_{cb} , discussed in section 6.2.1
- Functions of the height errors of tanks one, two and three, dH_i
- An indicator on where the change in throughput capacity is to be activated, TC .

These inputs are used to create set points for the regulatory controllers of sub-plants one to four.

Input 1, *InputFlow*: The first input to the controller is the *InputFlow* and this is the value of the maximum input into the total plant system that the source can provide. The range is [0 1] and represents a percentage of input flow (0 – 100%). This input is not used in the rule base as a condition, but in calculating the output. The outputs are functions of this input.

Input 2, SP_{cb} : This input, the cut-back set point, is also used to create the outputs. It has a range of $[0 \ 1]$, which represents a set point for the flow rate in percentage (%). SP_{cb} is calculated in the first part of the supervisory controller and explained in section 6.2.1.

Inputs 3, 4 and 5, dH_i : The next three inputs are functions of the height deviations of the three tanks with $dH_i = (H_i - 50)$ and $i = 1, 2, 3$. The value of dH_i will indicate how much the actual height differs from the nominal height of 50%. The height can vary between 40% and 60% before action is needed. Therefore, the membership function In , has a value of 1 in the range $[-0.2 \ 0.2]$, which represents $0.2 \times 50\% = 10\%$ to each side of the nominal 50%. This means that in this range (40% - 60%), the input is an absolute member of the function In . The membership function is a trapezoid, starting at -0.4 and ending at 0.4 , which is 20% from the nominal value and therefore an input smaller than 30% or greater than 70%, is not at all a member of In . If the height is not within boundaries (In), it is out of boundaries and can be either *Out Positive* or *Out Negative*. These functions overlap the membership function In in ranges $[-0.4 \ -0.2]$ and $[0.2 \ 0.4]$. Inputs smaller than -0.4 (30%), are absolute members of *Out Negative* and inputs greater than 0.4 (70%) are absolute members of *Out Positive*. These functions are shown in Figure 6.2.

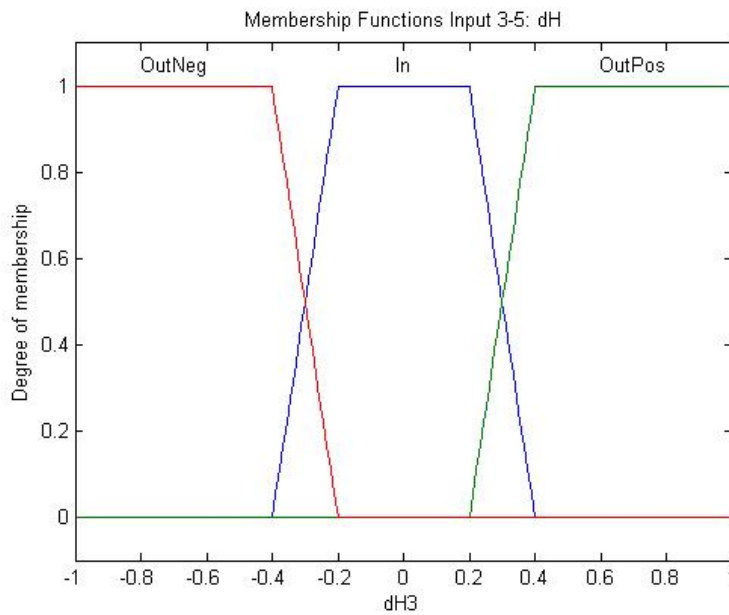


Figure 6.2 – Membership Functions For Inputs 3-5: dH_i

Input 6, TC : This input is the *Throughput Capacity* and indicates where a change in throughput capacity is activated. The operator will know beforehand when and where a capacity restriction or an increase in capacity will occur. This information is then entered into the controller through the input, TC . The value of the capacity restriction is used to determine the cut-back set point, SP_{cb} . This input, TC , only indicates where a restriction occur, despite the value of the restriction. A change in throughput capacity

can be due to capacity restrictions in either sub-plants one, three or four, or an increase or a decrease in the input flow rate of the whole system. The value of this input, TC , will indicate where a change in throughput capacity occurs at each time step. This input guides the decision making of which buffer to use in each situation. The possible values of the input and their meanings are shown in Table 6.1

Source of Change in Throughput Capacity	Value of TC	Description of Restriction
<i>Increase in Input</i>	0.9	Increase in maximum input
<i>Decrease in Input</i>	0.7	Capacity restriction or decrease in maximum input
<i>Sub-Plant 1</i>	0.5	Capacity restriction in sub-plant one
<i>Sub-Plant 3</i>	0.3	Capacity restriction in sub-plant three
<i>Sub-Plant 4</i>	0.1	Capacity restriction in sub-plant four

Table 6.1 – Capacity Restrictions And Changes In Throughput Capacity

The way this input is used can be explained by looking at the different situations:

Increase Input: When an increase in the maximum input occurs, the first sub-plant will automatically have the same throughput. The input into the first buffer will then be greater than the flow rate out of it and the height will increase. When the buffer's capacity is used to its limit and to prevent the tank from overflowing or to prevent dumping, the set point for the second sub-plant will follow the input. The same reasoning will apply for the next two buffers.

Decrease Input: A capacity restriction in the input means the flow into the first tank will be smaller than the throughput at that moment. The output valve from the tank can now be used to maintain the throughput and therefore keep the input, and therefore also the output of the next sub-plant, at the original throughput value. In a situation such as this, the flow into tank one is lower than the out flow from tank one. This will cause the height to drop. If the restriction in the input is only for a short period, the input flow to the tank will again increase to the original flow rate without any change in the flow rates through the rest of the sub-plants. If, however, the restriction in the input is not temporary, the output flow from the tank should be decreased when the height becomes too low. The output from the following sub-plant will then decrease and therefore also the input to tank two. Tank two will then be used exactly like tank one, keeping the output at the original value until the height becomes too low. The same will then happen to the third tank. If the capacity restriction lasted for a long period, the throughput will eventually decrease to the restricted value. If the restriction was valid only for a short period of time, the output from the plant system could have stayed at its original value.

Sub-Plant 1: For a capacity restriction in the first sub-plant, the same method is used, because this restriction is before the first buffer (tank one).

Sub-Plant 3: The next condition, where a capacity restriction occurs in the third sub-plant, is a bit different. The third buffer is used to maintain the output flow until the height of this buffer (tank three) becomes too low. The output is then decreased. Because of the fact that sub-plant three can only deliver a certain amount smaller than the original throughput, only that amount should be accepted from tank two. The flow rate out of tank two will then be less than the flow rate into the tank. Therefore the height will increase. This will happen until the level becomes too high and the input to tank two will then be decreased. The same will happen to the first tank, until the input is decreased. If the restriction is only for a short period of time, the output will not be affected. When a restriction occurs in the third sub-plant the time for which the original throughput can be maintained is much shorter, because there is only one buffer between the restriction and the output.

Sub-Plant 4: In the last case, a capacity restriction may occur in the last sub-plant. The output from the last sub-plant, and therefore the whole plant system, will have to decrease. The input to the sub-plant should be decreased to prevent dumping. The buffer capacity of the third tank will be utilised first. The height will increase because of the fact that the output flow is decreased and therefore smaller than the input. If the level becomes too high, the input flow must reduce and the output of the second tank should also decrease. This will increase its height. If the level is too high, the output from the first tank will reduce and after its maximum height limit is reached, the input flow and thus the total throughput will have to decrease.

In these situations, the buffers are used differently and that is why it is necessary for the controller to have information on exactly where the restriction is.

Outputs 1-4, $SetPoint1$, $SetPoint2$, $SetPoint3$, $SetPoint4$: A Sugeno type Fuzzy Logic controller is used for the supervisory control of the system. This means that the output membership functions are not Fuzzy variables, but functions of the inputs. All the outputs have the same membership functions and these membership functions are functions of inputs 1 and 2. The outputs from the supervisory controller are set point inputs to the regulatory flow controllers. The membership functions are SP_{mi} and SP_{cb} . The first membership function SP_{mi} is used when the height of the tank buffer is within boundaries and the set point is equal to the maximum input. The second membership functions is used when the buffer is used to its full capacity and a cut-back is activated. In this case the output set point is equal to the cut-back set point SP_{cb} . The rule base of the Fuzzy Logic controller is designed to generate a set point value for each regulatory controller by using the information of the changes in throughput capacity and the heights of the buffers. The Fuzzy Logic controller will weight all the rules and then set the output to either SP_{mi} or SP_{cb} or a combination of these two.

The first membership function, SP_{mi} is set equal to the first input, u_1 and the second membership function SP_{cb} is equal to the second input, u_2 .

Membership Function 1: $SP_{mi} = u_1$

Membership Function 2: $SP_{cb} = u_2$

A set of rules is used to determine each output. The output is a value equal to SP_{mi} or SP_{cb} or a combination of these two values. These rules are presented in Table 6.2. The rule base of the Fuzzy Logic controller are in the form of a set of *IF - THEN* rules. For example, the first rule states: *IF TC = Sub-Plant 4 THEN $SP_4 = SP_{cb}$* . This rule indicates that whenever a capacity restriction occurs in sub-plant four, the set point value for plant four should change to the cut-back set point.

The second rule is: *IF TC = Sub-Plant 4 AND $dH_3 = In$ THEN $SP_3 = SP_{mi}$* . If the capacity restriction occurs in sub-plant four, the buffer capacity of the third tank is used to decide whether the set point for the third plant should change. The second rule states that if the buffer is still in the safe range, the set point for sub-plant three is $SP_3 = SP_{mi}$ and it is not necessary to change the set point for the third sub-plant to the cut-back set point. If however the value of $dH_3 = OutPos$, the set point for the third sub-plant should change to the cut-back set point, SP_{cb} . This is stated in rule 4. The fuzzy input variables for height, will add a weighing function to the output and the output for the set point of sub-plant three, for example, can then be equal to SP_{mi} or SP_{cb} , or it can be a value in-between, depending on the fuzzy input dH_3 .

Normalization of the controller is necessary. The first two inputs are percentages (%) and should be in the range [0 1], therefore their fuzzification gains are equal to 0.01 ($FG_1 = FG_2 = 0.01$). The next three inputs (dH_i) are in the range [-50 50] and should be in [-1 1]. Therefore $FG_3 = FG_4 = FG_5 = 0.02$. The sixth input, TC , is entered in its fuzzy range [0 1], and normalization is not necessary. The outputs are all given as a value between 0 and 1 and should be a percentage (%). Therefore the defuzzification gains are all equal to 100 ($DG = 100$). Figure 6.3 shows a diagram of the supervisory controller.

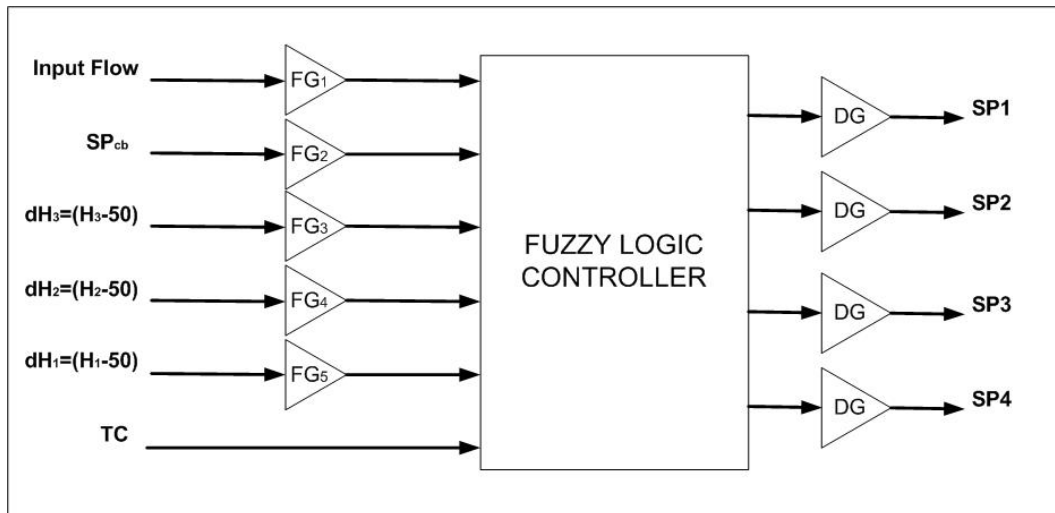


Figure 6.3 – Diagram Of Fuzzy Logic Controller For Supervisory Control

	Input 6 TC	Input 3 dH_3	Input 4 dH_2	Input 5 dH_1	Output 1 SP_4	Output 2 SP_3	Output 3 SP_2	Output 4 SP_1
1	Sub-Plant 4 (0.1)	-	-	-	SP_{cb}	-	-	-
2		In	-	-	-	SP_{mi}	-	-
3		Out Neg	-	-	-	SP_{mi}	-	-
4		Out Pos	-	-	-	SP_{cb}	-	-
5		-	In	-	-	-	SP_{mi}	-
6		-	Out Neg	-	-	-	SP_{mi}	-
7		-	Out Pos	-	-	-	SP_{cb}	-
8		-	-	In	-	-	-	SP_{mi}
9		-	-	Out Neg	-	-	-	SP_{mi}
10		-	-	Out Pos	-	-	-	SP_{cb}
11	Sub-Plant 3 (0.3)	-	-	-	-	SP_{cb}	-	-
12		In	-	-	SP_{mi}	-	-	-
13		Out Neg	-	-	SP_{cb}	-	-	-
14		Out Pos	-	-	SP_{mi}	-	-	-
15		-	In	-	-	-	SP_{mi}	-
16		-	Out Neg	-	-	-	SP_{mi}	-
17		-	Out Pos	-	-	-	SP_{cb}	-
18		-	-	In	-	-	-	SP_{mi}
19		-	-	Out Neg	-	-	-	SP_{mi}
20		-	-	Out Pos	-	-	-	SP_{cb}
21	Sub-Plant 1 (0.5) or Decrease In (0.7)	-	-	-	-	-	-	SP_{cb}
22		In	-	-	-	SP_{mi}	-	-
23		Out Neg	-	-	-	SP_{cb}	-	-
24		Out Pos	-	-	-	SP_{mi}	-	-
25		-	In	-	SP_{mi}	-	-	-
26		-	Out Neg	-	SP_{cb}	-	-	-
27		-	Out Pos	-	SP_{mi}	-	-	-
28		-	-	In	-	-	SP_{mi}	-
29		-	-	Out Neg	-	-	SP_{cb}	-
30		-	-	Out Pos	-	-	SP_{mi}	-
31	Increase In (0.9)	-	-	-	-	-	-	SP_{cb}
32		In	-	-	-	SP_{mi}	-	-
33		Out Neg	-	-	-	SP_{mi}	-	-
34		Out Pos	-	-	-	SP_{cb}	-	-
35		-	In	-	SP_{mi}	-	-	-
36		-	Out Neg	-	SP_{mi}	-	-	-
37		-	Out Pos	-	SP_{cb}	-	-	-
38		-	-	In	-	-	SP_{mi}	-
39		-	-	Out Neg	-	-	SP_{mi}	-
40		-	-	Out Pos	-	-	SP_{cb}	-

Table 6.2 – Fuzzy Logic Rules For Supervisory Controller

The supervisory controller is implemented on the cascaded system. The set points for the regulatory controllers are determined and sent to the different regulatory controllers, separately. The results are shown and discussed in section 6.3.

6.3 Experimental Results

The supervisory controller is implemented on the different regulatory controllers and evaluated. In this section the different methods are compared. A comparison between the regulatory controllers is done by looking at the output flow rates of each sub-plant and the heights of the different buffer levels in various situations.

6.3.1 Graphical Evaluation Of Time Responses With Supervisory Control

The supervisory control is tested with both the Fuzzy Logic controllers and the PI controllers. Although the output flow rates are similar with the PI and Fuzzy Logic control, the Fuzzy Logic controllers keep the heights within boundaries without dumping any products. The first test is done with PI control on sub-plants one to four, then with PI controllers on sub-plants one, two and four and MRAC on sub-plant three. It is done with Fuzzy Logic controllers on all the sub-plants as well. The supervisory controller sends the correct set point values to the regulatory controllers and they control each sub-plant. This is shown in section 6.3.1.1.

From these results, it becomes clear that the outputs of the sub-plants with Fuzzy Logic regulatory controllers are better because the height is taken into account in the controllers and changes are made just in time to keep the height within boundaries and still use the maximum buffer capacity. The supervisory controller is tested in different situations and the results from the Fuzzy Logic regulatory controllers are shown to confirm that they can take care of capacity restrictions in different areas of the plant.

6.3.1.1 Capacity Restriction In Sub-Plant 4 And In Maximum Input

For this test a capacity restriction occurs in sub-plant four after 1000 sec. Then after 6000 sec an additional capacity restriction occurs on the maximum input. If sub-plant three was linear, not much would have changed if a second capacity restriction of the same value occurred, but because it is not linear, the following can be seen in the figures: When the capacity restriction of 40% occurs in sub-plant four, the outputs of sub-plants three and four are 40% after a while, but the outputs of sub-plants one and two are greater than 40%, because the gain in sub-plant three is smaller than 1. If a capacity restriction of 40% then occurs on the maximum input, the outputs of sub-plants one and two should decrease to 40% and the flow rates after the non-linear sub-plant three will be less than 40%.

These results can be seen in the outputs of the PI regulatory controllers and Fuzzy Logic regulatory controllers in Figures 6.4 and 6.5. The outputs from the PI controllers with MRAC added to the third sub-plant are similar to the outputs from the PI controllers on all four sub-plants. The set point changes for the different types of controllers are similar. The set points for all four sub-plants are shown in Figure 6.6 for PI control and in Figure 6.7 for Fuzzy Logic control. The main difference is the heights of the liquid in the tanks. The results from the Fuzzy Logic regulatory controllers are better, because the heights are already taken into account and therefore the heights of the levels in the tanks did not exceed 80% or drop below 20% when using the Fuzzy Logic

regulatory controllers. With the PI and PI plus MRAC regulatory controllers, the valves are opened to their maximum and still cannot keep the height between 20% and 80%. The heights are shown in Figure 6.8 for PI control. PI controllers with a MRAC on sub-plant three show results similar to the PI controllers alone, but the heights from the Fuzzy Logic controllers are kept within the height boundaries and are shown in Figure 6.9. The heights of the levels in the tanks are evaluated throughout the different cases and the RMS error values for the heights are determined in section 6.3.2 and the different controllers are compared.

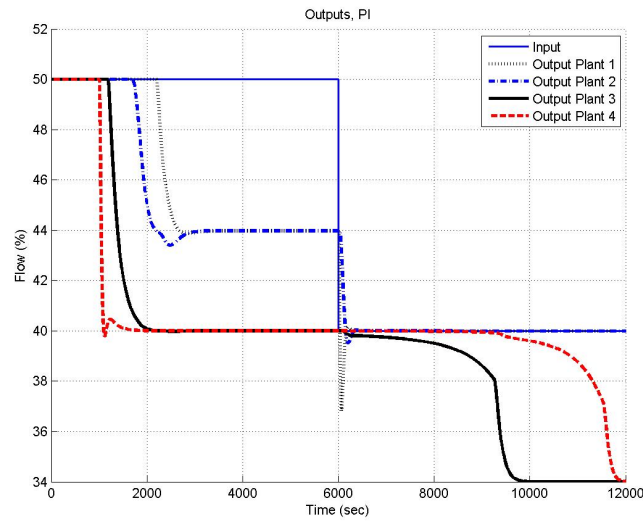


Figure 6.4 – Outputs From Different Sub-Plants With Capacity Restriction In Sub-Plant 4 And Then In The Maximum Input (Supervisory Control With PI Regulatory Control)

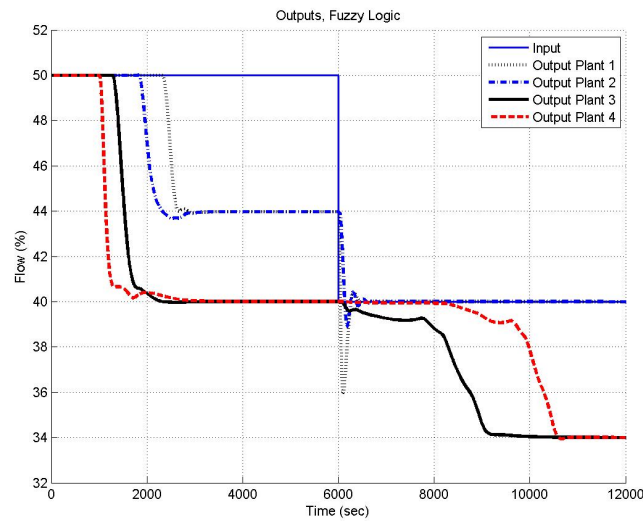


Figure 6.5 – Outputs From Different Sub-Plants With Capacity Restriction In Sub-Plant 4 And Then In The Maximum Input (Supervisory Control With Fuzzy Regulatory Control)

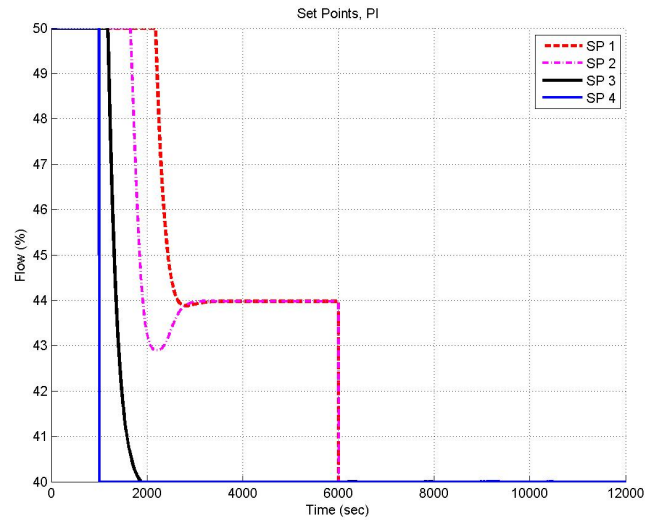


Figure 6.6 – Set Points For Regulatory Controllers With Capacity Restriction In Sub-Plant 4 And Then In The Maximum Input (Supervisory Control With PI Regulatory Control)

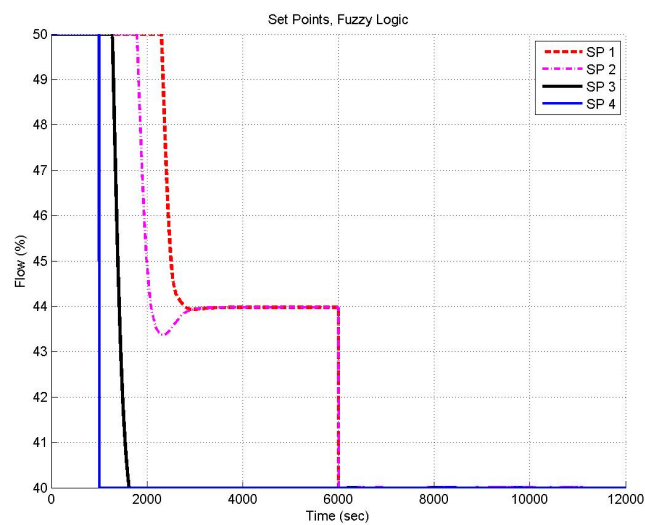


Figure 6.7 – Set Points For Regulatory Controllers With Capacity Restriction In Sub-Plant 4 And Then In The Maximum Input (Supervisory Control With Fuzzy Regulatory Control)

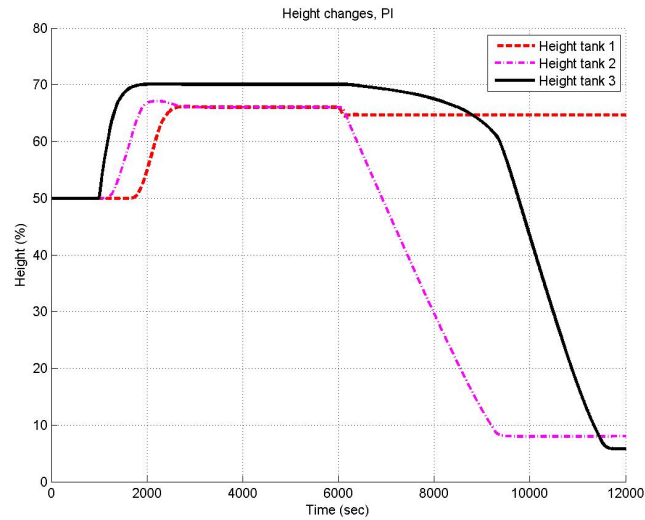


Figure 6.8 – Heights Of Different Tanks With Capacity Restriction In Sub-Plant 4 And Then In The Maximum Input (Supervisory Control With PI Regulatory Control)

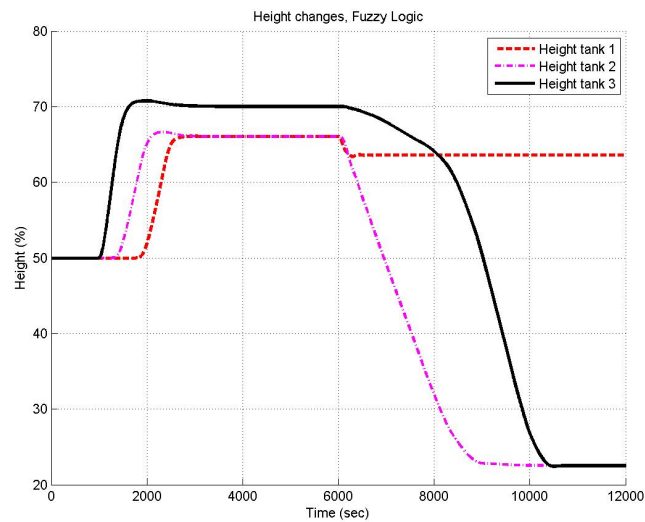


Figure 6.9 – Heights Of Different Tanks With Capacity Restriction In Sub-Plant 4 And Then In The Maximum Input (Supervisory Control With Fuzzy Regulatory Control)

Only the Fuzzy Logic regulatory controllers could keep the heights between 20% and 80% during all the simulation tests and the results are shown for the Fuzzy Logic regulatory controllers in Figures 6.10 to 6.16 in the next sections.

6.3.1.2 Capacity Restriction In Sub-Plant 1

After 1000sec a capacity restriction occurred in sub-plant 1.

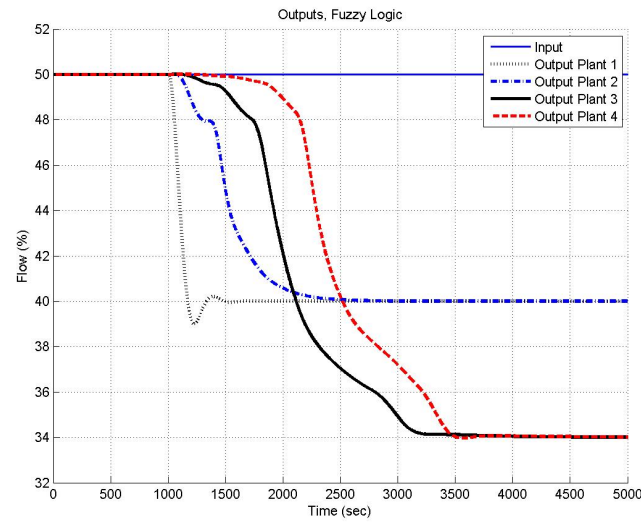


Figure 6.10 – Outputs From Different Sub-Plants With Capacity Restriction In Sub-Plant 1 (Supervisory Control With Fuzzy Regulatory Control)

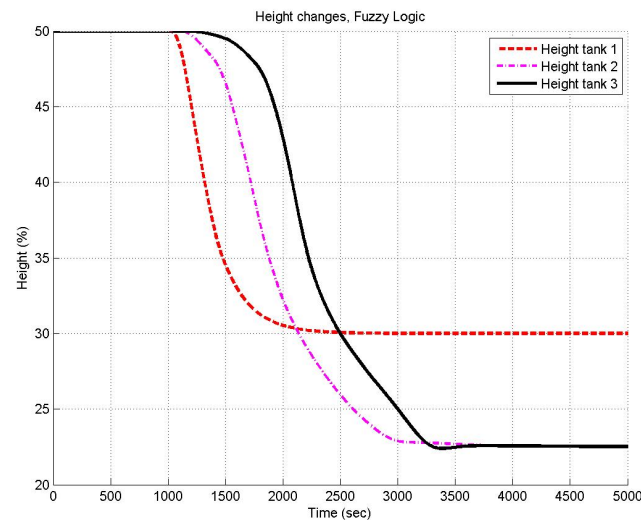


Figure 6.11 – Heights Of Different Tanks With Capacity Restriction In Sub-Plant 1 (Supervisory Control With Fuzzy Regulatory Control)

A capacity restriction in sub-plant one forces the output of sub-plant one to 40% after 1000sec. The height in tank one starts to drop, see Figure 6.11. Next, the output of sub-plant two drops to 40%, when the buffer capacity of tank one is used to its limit. Therefore the height in tank two drops. This forces the flow rate into sub-plant three to

drop and the outflow from sub-plant three will decrease even more, because it operates at a flow rate less than 50% and the gain of the third sub-plant is non-linear and smaller than one. As soon as the height in the last tank drops too low, the output from the last sub-plant drops to a flow rate equal to the flow rate from the non-linear sub-plant three, $Q = 34\%$.

6.3.1.3 Capacity Restriction In Sub-Plant 3

A capacity restriction occurs in sub-plant 3 after 1000 sec.

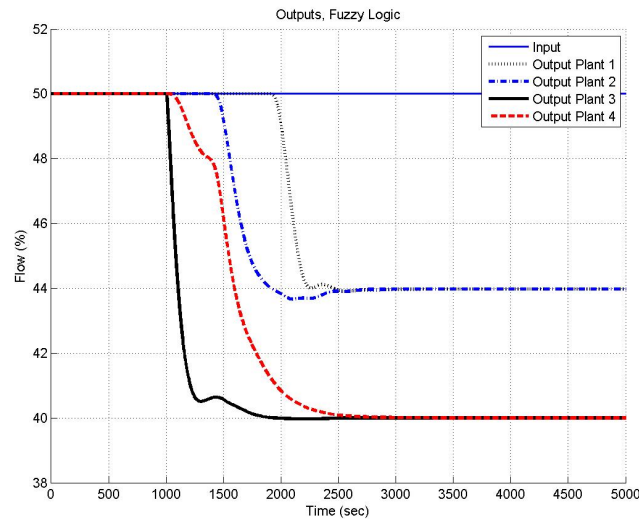


Figure 6.12 – Outputs From Different Sub-Plants With Capacity Restriction In Sub-Plant 3 (Supervisory Control With Fuzzy Regulatory Control)

The capacity restriction in sub-plant three forces the output of sub-plant three to 40%. As soon as the height in tank three drops too low, the output of sub-plant four follows to 40%. The height of tank two will increase because of the fact that the input flow rate is greater than the output flow rate. With the height of the liquid in tank two increasing, the input flow rate to tank two is decreased to prevent overflow. This is done by reducing the output flow rate of sub-plant two. This reduction causes the height in tank one to increase up to a certain point where it delivers the correct flow rate to keep the non-linear sub-plant three's output flow rate at 40%. This flow rate from sub-plants one and two are 44%.

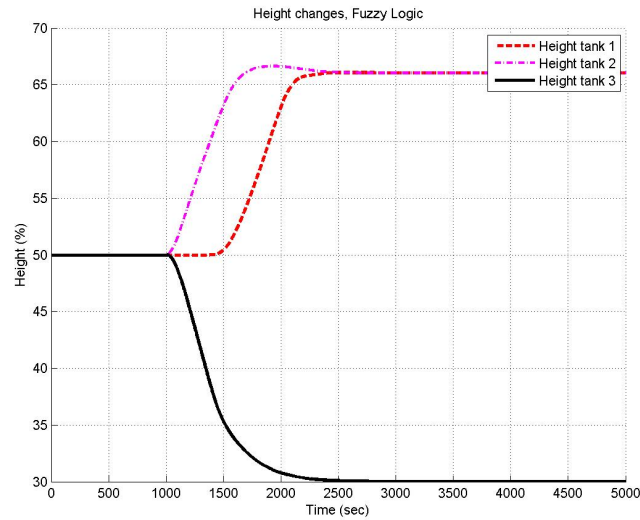


Figure 6.13 – Heights Of Different Tanks With Capacity Restriction In Sub-Plant 3 (Supervisory Control With Fuzzy Regulatory Control)

6.3.1.4 Capacity Restriction In Maximum Input For A Short Period

A capacity restriction occurs in the maximum input. This restriction occurs for a short period only. It started at 500 sec and ended at 1000 sec.

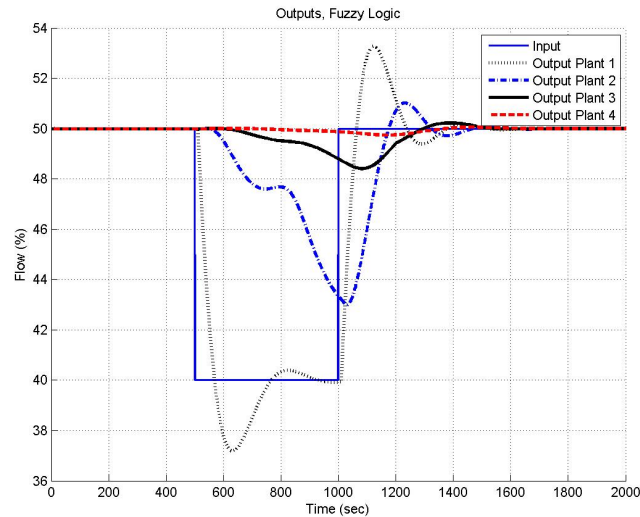


Figure 6.14 – Outputs From Different Sub-Plants With Capacity Restriction In Maximum Input For Only 500 sec (Supervisory Control With Fuzzy Regulatory Control)

When the maximum input has a capacity restriction for a short period of 500 sec, before returning to the nominal value of 50%, the output from tank one follows immediately. The first buffer is used to maintain the output from the second sub-plant, but a decrease

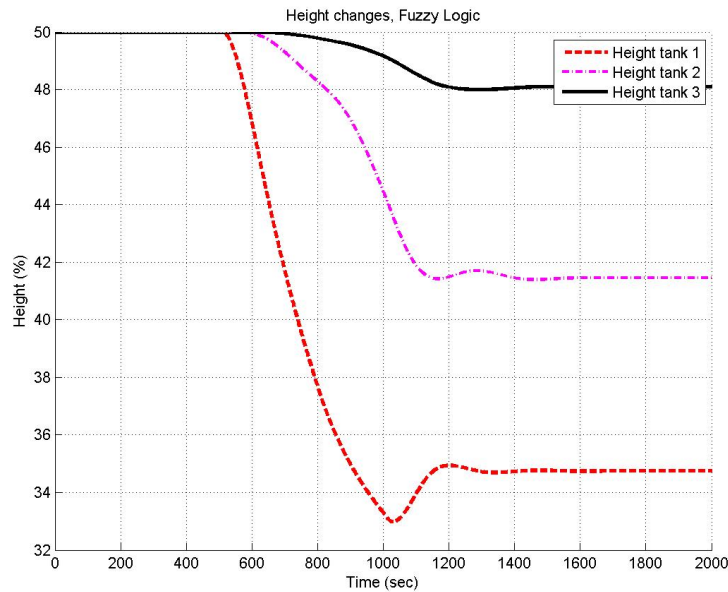


Figure 6.15 – Heights Of Different Tanks With Capacity Restriction In Maximum Input For Only 500 sec (Supervisory Control)

in the output from the second sub-plant is visible when the buffer capacity of the first buffer is used to its maximum capacity. Figure 6.15 shows the height values for all the buffers. The second buffer could almost keep the output flow from sub-plant three at 50%. Only a small deviation occurred. The third buffer could keep the output flow rate from sub-plant four and therefore the output flow rate from the whole plant system at 50%. The temporary capacity restriction in the input could therefore be absorbed throughout the buffered system, to keep the output steady at the nominal flow rate. The response of the output from sub-plant one shows a bigger overshoot than the overshoot for which it was designed in chapter 5. The reason for the big overshoot is that not only the set-point, but also the input into the plant changes at the same time. The set-point is changed when the maximum input experience a capacity restriction.

6.3.1.5 Increase In Maximum Input

The following graphs show the output flow and height results for an increase in maximum input at 1000 sec. No capacity restrictions occurred elsewhere. The first sub-plant will immediately follow the increase in maximum input, but the rest of the sub-plants will use the buffer capacities for as long as possible. This will ensure a minimum deviation in output flow. If the increase in maximum input is just temporary and for a short enough period, the output of the fourth tank will remain unchanged. The fact that the set point of the first sub-plant increases at the same time that the input into the sub-plant increases, enlarge the maximum overshoot of the closed loop response of the first sub-plant.

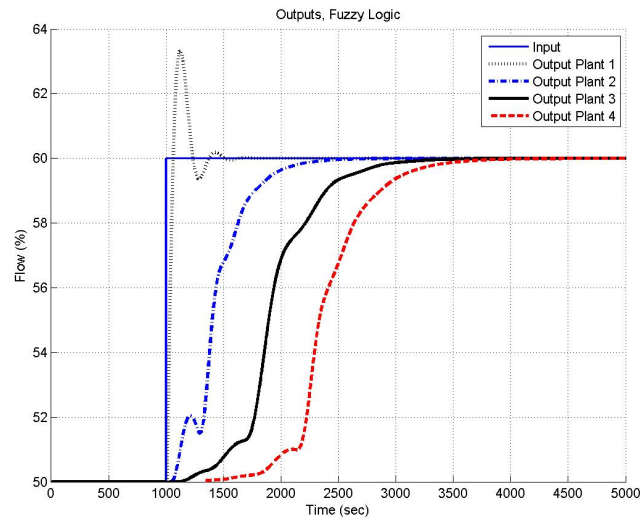


Figure 6.16 – Outputs From Different Sub-Plants With Increase In Maximum Input After 1000 sec (Supervisory Control With Fuzzy Regulatory Control)

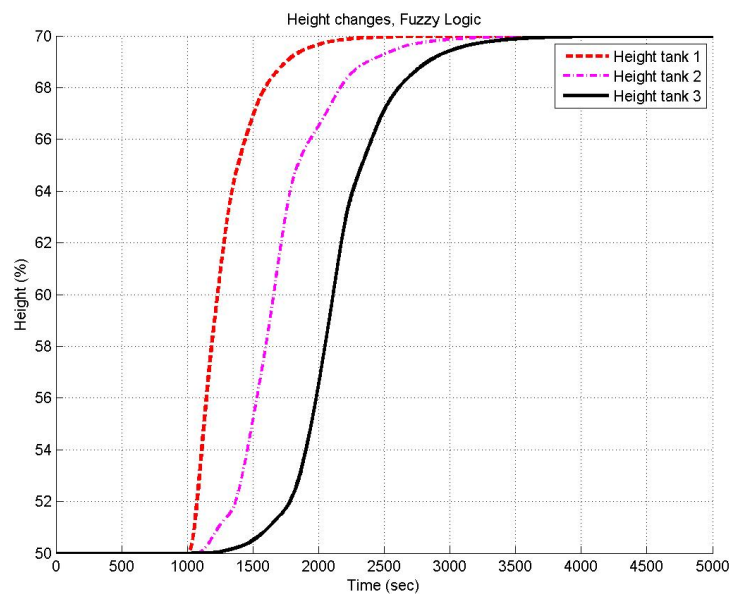


Figure 6.17 – Heights Of Different Tanks With Increase In Maximum Input After 1000 sec (Supervisory Control With Fuzzy Regulatory Control)

6.3.2 Statistical Evaluation Of The Height errors During Supervisory Control

To evaluate the controllers statistically, the same RMS value, used in section 5.4.3, is used to determine which controller kept the heights of the tanks inside of the preferable range (40% - 60%) and inside the safety range (20% - 80%). The RMS values were determined twice. Once where the error signal is the value of the heights outside a 10% range (smaller than 40% or greater than 60%). The second error signal is created where the height is outside the 30% range (smaller than 20% or greater than 80%). The RMS values are shown in table 6.3. The same tests as in section 6.3.1 are done. For the 30% range RMS values, only the tests where the heights are outside the boundaries are shown.

	Height 1	Height 2	Height 3
Control Method	10% RMS Error	10% RMS Error	10% RMS Error
CR in 4, then in Input			
PI	5.1	3.61	13.31
PI+MRAC	5.166	3.62	12.98
FUZZY	4.2	3.6	10.15
CR in 1			
PI	8.2	20.6	17.93
PI+MRAC	8.25	20.73	17.778
FUZZY	8.16	12.78	11.82
CR in 3			
PI	4.56	5.01	8.09
PI+MRAC	4.59	5.05	8.15
FUZZY	2.72	3.449	5.2144
Increase In Input			
PI	8.29	7.61	7.73
PI+MRAC	8.3	7.6	7.76
FUZZY	8.28	7.59	7.73
	30% RMS Error	30% RMS Error	30% RMS Error
CR in 4, then in Input			
PI	0	0	3.543
PI+MRAC	0	0	3.64
FUZZY	0	0	0
CR in 1			
PI	0	0	6.03
PI+MRAC	0	0	5.99
FUZZY	0	0	0

Table 6.3 – Root-Mean-Square Errors

It can be seen that the Fuzzy Logic controllers have the smallest RMS value outside the 10% borders and the heights are never outside the 30% safety boundaries. With PI and MRAC, the heights are not always kept within 30% boundaries.

6.4 Conclusion

The Supervisory Control for the whole plant system with four sub-plants in series is tested and results are shown in section 6.3.

The supervisory controller shows the best results with the Fuzzy Logic Controllers on all four sub-plants. The reason for this is that the Fuzzy Logic controller is the only regulatory controller that uses the height as well as the flow rate to calculate the different valve positions. The PI and MRAC controllers use only the flow rate to determine the valve positions for the different sub-plants. As a result, the heights are not kept within the restrictions set for the liquid level.

With the Fuzzy Logic controllers the height restrictions were honoured and changes were made smoothly. Although the PI control plus MRAC on the third sub-plant showed better performance of the disturbance rejection than the Fuzzy Logic control, it did not honour the limits of the buffers. The results from the Fuzzy Logic controllers on all four sub-plants are still acceptable and disturbances were rejected satisfactorily.

Fuzzy Logic control is therefore proposed to be used in both control layers: Regulatory as well as Supervisory control.

Chapter 7

Conclusion

In this thesis, different control strategies to control the cascaded plants with buffering have been investigated. The goals were to minimise the influence of process disturbances and to maximise the process yield at the output. The control of the whole plant system was divided into two different control layers. The inner layer consisted of the regulatory controllers for each sub-plant individually, and the outer layer was the supervisory controller of the whole plant.

This chapter aims to conclude the work that has been done. Firstly the measurement criteria will be summarised in section 7.1. A comparison between the three main control techniques used, will be covered in section 7.2. After this comparison, a concluding summary of the different results will be presented in section 7.3. In the last section (7.4) a few recommendations for further research are offered.

7.1 Measurement Criteria

The measurement criteria, described in section 1.1.3, were used to compare the different sets of controllers and to draw conclusions. As stipulated in that section, base line control was designed first, and then other control methods were compared to the base line control. For base line control, PI control was used. The other control methods are Fuzzy Logic and Model Reference Adaptive Control. The comparison was done by adding different disturbances, described in section 1.1.4, to the system and then comparing the control methods. The control methods were compared with a graphical analysis as well as a statistical analysis of the flow rates throughout the whole plant system, the frequencies of disturbances and the heights of the liquid in the buffers. The results from these measurements were discussed at the end of each chapter and a concluding summary of the results will be given in section 7.3.

The main control objectives were described in section 1.1.2, and were three-fold. The first was to maximise the throughput and the total process yield. The second objective was to minimise the loss of product through dumping valves. The combination of these two maximised the profit. The third control objective was to honour the limits of the buffers. All the controllers will be measured against these objectives in sections 7.2 and 7.3.

7.2 Controller Comparison

To compare the different controllers, a summary of the advantages and disadvantages of each controller covered in chapters 3 to 6, are given below:

For the base line control of the first simplified model with two tanks, Linearised State Feedback control was used. This control method was very effective. The controller was designed for a linearised system at a certain work point. At this work point the controller performed the best, but control at other work points was sufficient. This controller had a rather fast response and disturbances were rejected well.

For the height and flow base line control of the more realistic representation, PI control was used. With PI control good disturbance rejection could be accomplished. The set point changes were followed well. The PI controllers showed faster responses than the Fuzzy Logic controllers for flow control and even the non-linear sub-plant could be controlled at values away from the nominal values where the controllers were linearised. The performance of the non-linear sub-plant did regress when it operated at values where $Q < 0.01m^3/sec$, though. Multi variable control was more difficult and not feasible when an accurate mathematical model was not available. The fact that only the flow rate was used to control the valve position resulted in insufficient height control and buffer limits were exceeded.

Model Reference Adaptive Control worked well at flow rates different from the nominal flow rate for flow and height control. The control parameters adapted to new values when the system changed. A disadvantage is the fact that for multi-variable control, an accurate mathematical model is needed. The heights were not taken into account when determining the valve positions and therefore the height limits were exceeded at times.

Fuzzy Logic height control was very slow without the addition of derivative action. When derivative action was added, the Fuzzy controllers were fast and sufficient for following set point changes and rejecting disturbances. The addition of derivative action can increase the noise if the system is noisy, though. Fuzzy Logic flow control was slower than the other two methods for the regulatory flow control, yet the disturbance rejection was still good and set points were followed fast enough. No mathematical model was necessary for multi-variable control. The fact that fuzzy variables are used, made it easy for the operator to understand and adjust the controller. The controller performance was good at values away from the nominal values. Also, the non-linear sub-plant's controller performed just as well at a flow rate lower than 50%. The fact that multi-variable control could be implemented without a mathematical model, made it possible to honour the buffer limits under different circumstances.

7.3 Conclusion Of Results

In this thesis four different steps of control can be distinguished. The first step was the control of the simplified model. Here, the flow rate and temperature control for a two-tank system was designed. The second step was the control of the height of the liquid in the buffer. From here, the third step was taken where the flow rates of each individual

sub-plant were determined. The last step was to implement supervisory control on top of the regulatory flow control. In conclusion, the results from all these different steps are described next.

The first simplified model was a two tank system, controlled by three different types of controllers, as described in chapter 3. Set point changes in flow rate or temperature had similar responses with the Linearised, Fuzzy Logic and Model Reference Adaptive Controllers. For flow control, disturbance rejection was the best done by the Fuzzy Logic controllers, while the linearised controllers showed the best results in disturbance rejection for temperature control. The Model Reference Adaptive Controllers followed set point changes well and rejected temporary disturbances, but disturbances in the input flow and input temperature had a greater effect on the outputs than when using the Fuzzy Logic or Linearised controllers.

A more realistic representation of the system was introduced in chapter 4, where height control was done to control the buffers. Here the MRAC and Fuzzy Logic controller showed the best results, with the heights not deviating much from the set point value of 50% even with disturbances entering the system. All the controllers acted sufficiently in following set point changes. For sufficient disturbance rejection in flow rate, though, the buffers need to be used to absorb the disturbances, and height control was not the best control approach.

Regulatory flow control (chapter 5) was the next control approach investigated. PI controllers, a Model Reference Adaptive Controller for sub-plant three and Fuzzy Logic controllers were designed. These different types of controllers were compared and the PI controllers rejected disturbances the best. The addition of a MRAC on sub-plant three made disturbance rejection slightly worse, but the results were rather similar. With the Fuzzy Logic controllers, the outputs deviated a bit more from the set point values when disturbances entered the system, but the disturbances were still rejected satisfactorily. The Fuzzy Logic regulatory flow controllers not only used the flow rates from the sub-plants to determine the valve positions, but also took the current height into account, which made it possible to use the buffers to their full capacity and still keep the buffer height within boundaries. This increased the safety, because the storage tanks would not run empty or overflow.

In chapter 6 the design of a supervisory controller was the next step to investigate. The supervisory controller was designed to replace the human operator, by being able to make decisions and to regulate the total throughput. This was done to optimise the total profit by minimising losses and maximising total process yield. The supervisory controller also kept the flow rates from the system constant or changed them smoothly, by absorbing frequent small changes in the buffering through the system. Multi variable control was made easy by Fuzzy Logic control. No mathematical models were necessary to control the overall system. Fuzzy variables that were used to describe the controlled variables are everyday terms and can easily be understood by the operator. The supervisory controller automated the factory throughput. Set point changes for the sub-plants were determined by the supervisory controller and the regulatory controller for each sub-plant followed its own set point. Because of the fact that the Fuzzy Logic controllers made it possible to control the flow rate and still keep the heights between

the allowable values, the Fuzzy Logic regulatory controllers performed the best with the Fuzzy Logic supervisory controller.

The use of both control layers, regulatory and supervisory, increased the total throughput and prevented losses. This was done while the equipment was kept within safe limits. The control of the cascaded plant system with buffering could therefore be implemented to improve total plant yield.

7.4 Further Research And Recommendations

There are limited publications available on the topic investigated in this thesis, i.e. the control of cascaded plants with buffering. This topic was investigated and the goals was to minimise the influence of process disturbances and to maximise the process yield at the output. These goals were reached, however further research on this topic are still required.

The coal to liquid facility, aimed for in this thesis, consists of two identical, parallel process trains, as described in chapter 1. In future, both parallel trains can be examined and controlled, to reach the control objectives. In this thesis, only one of the two process trains was controlled. When both trains are considered, capacity restrictions that occur in only one of the trains can be overcome by receiving the limited substance from the other train, before cut backs are activated. This can increase the total process yield, without losing product through dumping or flaring.

The control of the cascaded plant system was only simulated, and although the tank models reacted as expected, the control was not implemented on an actual system. In future, the control methods proposed could be tested on a physical plant.

The general field of supervisory control of cascaded systems in the process control industry is very important and can be used to further optimise the total closed loop process performance.

Bibliography

- [1] Åström, K. and Wittenmark, B.: *Adaptive Control*. 2nd edn. Addison-Wesley Publishing Company, 1995.
- [2] Foss, B.A. and Halvorsen, I.J.: Dynamic optimization of the liquefied natural gas value chain. Annual Gas Processing Symposium, Department Of Engineering Cybernetics N-7491 Trondheim, Norway, 2009.
- [3] Strydom, J.: Coal to liquid production facility. Personal Communication, 2010. Sasol Secunda Factory.
- [4] Dorf, R. and Bishop, R.: *Modern Control Systems*. 10th edn. Pearson Prentice Hall, 2005.
- [5] Jantzen, J.: Fuzzy control. Lecture Notes in On-Line Process Control (5354), Technical University of Denmark, October 1991.
- [6] Gopal, M.: *Digital Control and State Variable Methods*. 2nd edn. McGraw-Hill, 2004.
- [7] Kaya, I., Tan, N. and Atherton, D.: Improved cascade control structure for enhanced performance. *Journal of Process Control*, vol. 17, pp. 3–16, 2007.
- [8] Lestage, R., Pomerleau, A. and Desbiens, A.: Improved constrained cascade control for parallel processes. *Control Engineering Practice*, vol. 7, pp. 969–974, 1999.
- [9] Jantzen, J.: Fuzzy supervisory control. Tech. Rep. 98-H-875, Technical University of Denmark, Department of Automation, November 1998.
- [10] Castro, J. and Doyle, F.: A pulp mill benchmark problem for control: Application of plantwide control design. *Journal of Process Control*, vol. 14, pp. 329–347, 2004.
- [11] Skogestad, S.: Plantwide control: The search for the self-optimizing control structure. *Journal of Process Control*, vol. 10, pp. 487–507, 2000.
- [12] Radhakrishnan, V.: Model based supervisory control of a ball mill grinding circuit. *Journal of Process Control*, vol. 9, pp. 195–211, 1999.
- [13] Tosukhowong, T., Lee, J., Lee, J. and Lu, J.: An introduction to a dynamic plant-wide optimization strategy for an integrated plant. *Computers and Chemical Engineering*, vol. 29, pp. 199–208, 2004.
- [14] Xu, S. and Bao, J.: Distributed control of plantwide chemical processes. *Journal of Process Control*, vol. 19, pp. 1671–1687, 2009.

- [15] Bhattacharya, S. and Bose, S.: Mathematical model for scheduling operations in cascaded continuous processing units. *European Journal of Operational Research*, vol. 182, pp. 1–14, 2007.
- [16] Cheung, T. and Luyben, W.: Liquid-level control in single tanks and cascades of tanks with proportional-only and proportional-integral feedback controllers. *Industrial & Engineering Chemistry Fundamentals*, vol. 18, no. 1, pp. 15–21, 1979.
- [17] Mercangöz, M. and Doyle, F.: Distributed model predictive control of an experimental four-tank system. *Journal of Process Control*, vol. 17, pp. 297–308, 2007.
- [18] Engell, S.: Feedback control for optimal process operation. *Journal of Process Control*, vol. 17, pp. 203–219, 2007.
- [19] Faanes, A. and Skogestad, S.: Controller design for serial processes. *Journal of Process Control*, vol. 15, pp. 259–271, 2005.
- [20] Hay, G., Svrcek, W., Ross, T. and Young, B.: Control loop noise rejection using fuzzy logic. *ISA Transactions*, vol. 44, pp. 457–464, 2005.
- [21] Ismail, R. and Jusoff, K.: Fuzzy state space model of multivariable control systems. *Computer and Information Science*, vol. 2, no. 2, pp. 19–25, 2009.
- [22] Shen, S. and Lee, T.: Synthesis of uncertain cascaded multiple loops systems with time delay. *Journal of the Franklin Institute*, vol. 335B, no. 4, pp. 789–798, March 1997.
- [23] Gopal, M.: *Control Systems*. 2nd edn. McGraw-Hill, 2003.